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Worley
Edmonton, AB

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T6G2R3

April 8th, 2020

Dear Mr. Nolte and Mr. De Clerk,

On behalf of TR Solution, we are pleased to submit the preliminary design proposal for the construction of a standalone slurry hydrocracker (SHC) unit as a potential oil sands bitumen upgrading facility in Edmonton. This report was prepared to evaluate the economic feasibility of the process.

The proposed SHC facility is designed to convert 95% of 100,000 barrels per day of hot bitumen feed and includes a diluent recovery unit (DRU), a steam methane reformer (SMR) and pressure swing adsorption (PSA), a slurry bubble column reactor (for combined bitumen hydro-conversion and thermal cracking), high pressure separation units, a stream-stripping process, an amine sweetening process, and a sulfur recovery unit. The project design contains detailed design specifications for process equipment, process flow diagrams, environmental, safety and economic analyses. The completion of the proposed plant including basis of design, engineering, procurement, construction and commissioning phases are estimated to have 5 years completion time. Therefore, the facility has June of 2025 as the proposed start-up date, and it requires an initial capital investment of 6.9 million (\$CND, Edmonton, 2020).

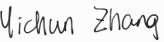
TR Solutions would like to thank Mr. Nolte and Mr. De Clerk for all the help, feedback and continuous support during the development of the project.

Sincerely,

Naira Correia 

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SLURRY HYDROCRACKER PROJECT

Final Report

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EXECUTIVE SUMMARY

This report outlines the preliminary design information and plan for a proposed greenfield slurry hydrocracker (SHC) facility based in Edmonton, Alberta. This facility is a partial bitumen upgrader designed for a capacity of 100,000 barrels per day of hot bitumen feed from a steam assisted gravity drainage producer. Bitumen upgrading is widely used in Alberta to refine heavier oils to more useful crude derivatives. SHC technology is an emerging alternative to conventional methods, promising higher feed conversion and efficiency. The main business driver for this project is the continuously increasing crude oil demand globally and the increasing incentive of producing higher quality crude. The SHC process is expected to produce higher quality crude to meet increasing demand, presenting a potential profitable investment for Alberta's growing energy industry.

A detailed design of a battery limit slurry hydrocracker process is proposed to process 100,000 barrels equivalent of hot bitumen to produce 123,500 bbl/day of premium quality crude with Naphtha, Light Gas Oil and Heavy Gas Oil distributions of 21.9 wt.%, 40.2 wt.% and 35.7 wt.% respectively. The facility includes the auxiliary process facilities (hydrogen plant, amine treatment unit, Claus plant and diluent recovery unit) for a standalone site greenfield project. The proposed location of the project is in Edmonton, Alberta. VMG Symmetry was utilized to characterize feed and product oils using the Refinery Advanced Peng-Robinson as the thermodynamic model.

The project has an NPV of \$-4.68 Billion CAD by the end of the 35th operating year. The ROC is 0.10% and the payback period is 997 years. One of the reasons this project is not economically feasible is because the capital cost for the SMR unit is very high, with a \$589 Million CAD of capital, which is even higher than the project scope (hydrocracker) capital cost. In terms of operating cost, bitumen feed took 50% of the total manufacturing cost. The price of bitumen is high, and the margin between the price of bitumen and crude oil products is \$150.25 CAD/m³, which is not enough to generate a satisfying cash flow. A sensitivity analysis proved that the NPV is very sensitive to crude oil price and total capital cost. For the project to have a NPV of zero by the end of the cash flow, the crude oil should be priced at \$659.83 CAD/m³. One way to make the project more feasible is to have an altered option to purchase all the hydrogen from other sources

in Alberta. If all the hydrogen is purchased at \$3 CAD/ft³, the NPV will become \$-1.14 Billion CAD.

This project will not pose any significant environmental concerns, as the only form of emission is greenhouse gases (GHG). The Oil Sands Emissions Limit Act of Alberta outlines a maximum GHG emission from an upgrader facility of 10 megatons a year. The total direct and indirect GHG emission of the project will be 5.19 megatons per year. A CO₂ sequestration unit is a potential strategy to further reduce CO₂ emission as well as to generate additional profit.

A Process Hazard Analysis (PHA) was performed on the entire facility, and a hazard and operability (HAZOP) study was performed on the piping and instrumentation diagram (P&ID) of a separator unit. Potential risks associated with deviations in temperature, pressure, and liquid levels will be controlled and reduced to low or medium-low levels. It is recommended that all workers on site must wear personal protective equipment (PPE) all the time, and management must prepare accessible procedures and an evacuation plan. A plot plan was designed based on the explosion and chemical exposure radii, in which hazardous equipment were distanced from one another.

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1. INTRODUCTION & STATEMENT OF OBJECTIVE

This report outlines the scope and plan for the design and construction of a proposed standalone oil sands bitumen upgrading facility utilizing slurry hydrocracking in Edmonton, Alberta. Slurry hydrocracking is an upgrading technology that converts low-quality bitumen into lighter, more valuable hydrocarbon products. The proposed process achieves higher conversion and product yields than the existing upgrading facilities in Alberta that rely on carbon rejection technologies for upgrading.

The objective of this greenfield project is to develop the economic basis for constructing a standalone slurry hydrocracker unit that can process 100,000 bbl/day of bitumen feed. The economic driver of this project is the financial incentive of producing a high yield, high quality synthetic crude product through high feed conversion while complying with all applicable plant safety and environmental standards.

2. BACKGROUND

In the past 10 years, world oil demand has increased by 19% and is projected to increase another 12% from 2019 to 2040^[1,2]. Canada holds 10% of the world's reserves of crude oil, and 96% of Canada's crude oil production is based in oil sands extraction^[3]. The product of oil sands extraction is crude bitumen. Bitumen is a complex mixture of heavy hydrocarbons with sulfur, nitrogen, and heavy metal impurities that need to be processed to make usable products, such as fuels. This is achieved through bitumen upgrading, whereby heavy hydrocarbons are broken down into smaller, usable compounds.

The majority of bitumen upgrading performed in Alberta is achieved through carbon rejection processes, such as delayed coking or ebullated bed reactions. Compared with more modern and advanced methods of upgrading, carbon rejection methods are subject to low product yields and low feed conversions, as they produce an undesirable solid residue called coke. Recent developments in partial upgrading technologies, such as Slurry Hydrocracking (SHC), promise higher rates of conversion and yield^[4]. SHC is a process whereby a residue feed (in this case Athabasca bitumen) and a catalyst are mixed in an upflow reactor in a hydrogen-rich environment. The process includes both catalytic hydrogenation and thermal cracking of bitumen to achieve a high residue conversion into lighter products with zero coke formation. The process simultaneously

involves hydrogenation of alkenes and aromatics to increase product hydrogen-to-carbon ratio, and therefore quality^[4]. This technology may be a potential profitable investment for Alberta's energy industry. TR Solutions is evaluating the economic basis of building a SHC unit with 100,000 barrels per day (bbl/d) capacity as requested by Worley Limited.

3. PROPOSED SOLUTION

3.1 Overview

The technical design basis of this project is to develop a standalone bitumen upgrading facility utilizing SHC technology which can process 100,000 bbl/day of hot bitumen feed. The block flow diagram of the proposed solution is presented in Figure 1.

The technical design, sizing and costing using AACE Class IV estimates will be conducted on equipment in the hydroconversion reactors, heat exchangers and the downstream high-pressure separation vessels including two phase and three phase separators and stripper tray column. All required auxiliary utilities excluding electricity are supplied inside battery limits (ISBL). The OSBL utilities include a steam generator and a cooling tower to provide high pressure steam and cooling water respectively. Natural gas will be used as a fuel source and as a process stream for hydrogen production.

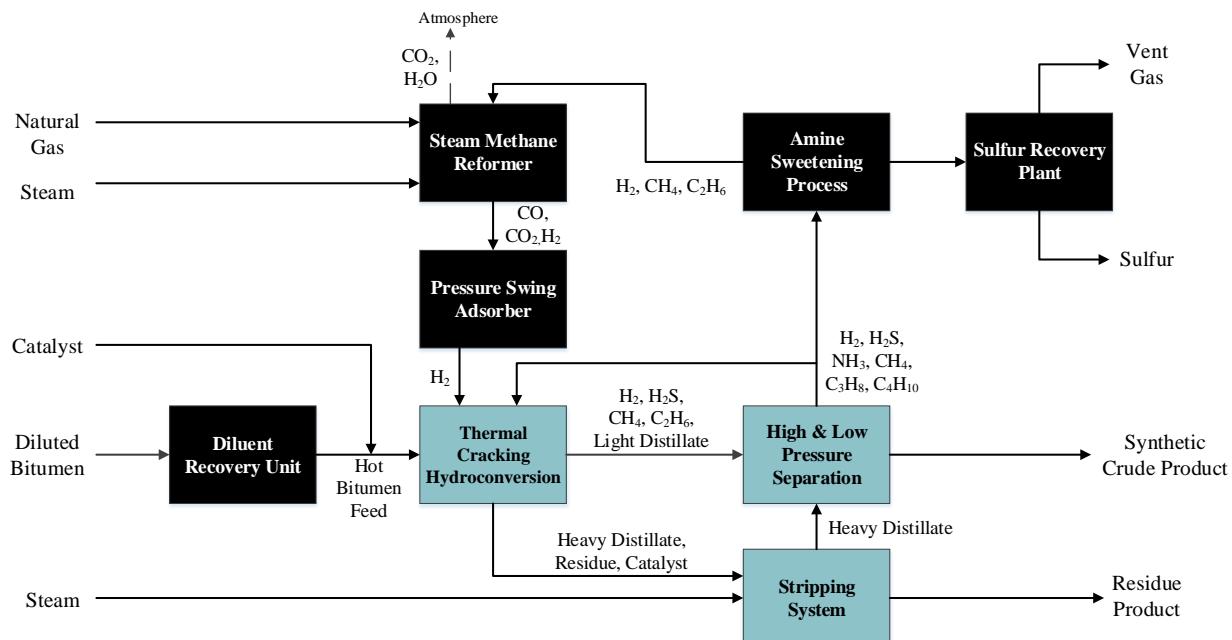


Figure 1: Block flow diagram for the proposed design.

3.2 Process Summary

The scope of the proposed design entails a diluent recovery unit (DRU), a steam methane reformer (SMR) and pressure swing adsorption (PSA) pair (for hydrogen generation and purification), a slurry bubble column reactor (for combined bitumen hydroconversion and thermal cracking), high and low pressure separation units, a stream-stripping process, an amine sweetening process, and a sulfur recovery unit.

The original feedstock for the proposed plant is assumed to be naphtha-diluted bitumen (dilbit), transported via pipeline to the facility from a steam assisted gravity drainage (SAGD) site. The dilbit is processed in a diluent recovery unit to separate the bitumen and naphtha diluent, whereby the bitumen can be further processed, and the diluent can be re-sold. The bitumen feed will be characterized as conventional Athabasca bitumen with a negligible amount of heavy metal impurities.

The DRU is a fractionation system that distils the naphtha diluent and extracts hot bitumen for processing. The recovered bitumen is heated and mixed with catalyst, and the slurry mixture is routed through the hydroconversion reactors. While this happens, the reactors are sparged with hydrogen-rich gas, and the solution undergoes hydrocracking reactions. These reactions convert the heavy residues present in bitumen to lighter hydrocarbon molecules. Byproducts of the reaction include the separation of sulfur and nitrogen heteroatoms to hydrogen sulfide (H_2S), ammonia (NH_3), and water. The specific type of reactor used are slurry-phase bubble column reactors.

The lightest, most valuable liquid product produced is released from the reactor as a vapour and is sent through a series of high and low pressure vapor-liquid separations. This separates out the value liquid hydrocarbons from non-condensable gases (primarily unreacted hydrogen, light saturated hydrocarbons, H_2S , and NH_3). The heavier, less valuable residue stream exits the reactor as a liquid, and contains unreacted feed, some heavy distillate, and the spent catalyst. This stream is routed through a stripping system to recover a low-residue, heavy distillate, and produce a residue product (containing unconverted feed and spent catalyst). The heavy distillate is combined with the light hydrocarbon liquid from the vapour-liquid separations and is put in a pipeline and sold. The residue product is given to a nearby coking unit (outside battery limits) where it can be further processed in valuable products.

The hydrogen rich overhead gas (recovered in the high pressure separation) is split into two streams; one “internal” recycle stream, which (after being re-compressed and re-heated) is sent

back into the reactor to maintain high hydrogen partial pressure, and one “external” recycle stream, which is sent to an amine sweetening unit to remove H₂S and NH₃. The sweet gas from the amine unit will be fed into the SMR as a reactant for hydrogen production. This will leverage the purge stream as a valuable process stream (instead of a fuel utility stream). In addition to this, light hydrocarbon overheads from the stripping system will be sent through the amine unit whereby they will be fed as SMR reactants as well.

The SMR combines this light hydrocarbon stream with natural gas make-up and mid-pressure steam to produce syngas (a mixture of hydrogen, carbon dioxide, and carbon monoxide) with a catalytic reactor. The syngas is sent to the PSA, which includes multiple pressurized bed columns, to separate hydrogen from the syngas mixture by adsorbing non-hydrogen components. This leaves a pure-hydrogen stream to be compressed and fed into the reactor where it will account for hydrogen consumption in the reactor, as well as various hydrogen losses in the system. The CO and H₂ rich adsorbed components are eventually desorbed from the PSA and are fired as fuel gas.

The scope of detailed design work to be completed is illustrated in Figure 1 by the colour of the process units. Detailed equipment design will not be performed on the ‘black boxes’, including the sulfur plant, amine process, DRU, SMR, and PSA units. Capital and operating cost for these units will be estimated and scaled up based on the required capacity and industrial cost references.

3.3 Business Case

This project aims to design a SHC bitumen upgrading system which promises higher overall conversion of bitumen feed than conventionally used delayed coking and ebullated bed reaction methods ^[4]. The main economic driver for this project is the incentive of selling a high volume of quality crude product at a high price.

One of the requirements for this project is to have a high conversion of feed bitumen. The typical conversions for ebullated bed reactor processes range from 60%-80%, whereas hydrocracking can achieve conversion greater than 90%, which will increase the quality of crude oil produced and profit margin ^[5].

In this scope, the crude product will be characterized as referenced by SHC pilot plant CANMET product and it will be priced based on three boiling point cuts; naphtha (50-170°C), light gas oil (170-350°C), and heavy gas oil (350-550°C) ^[6]. The product is priced as a discount or premium indexed to the Western Texas Intermediate (WTI) benchmark. The reference price for

the bitumen product and natural gas is the average price from January to December in 2019. A few factors that affected the price of the crude oil include, the density (usually defined by the American Petroleum Institute gravity scale (^oAPI)), hydrogen to carbon (H/C) ratio, and sulfur content. In general, low densities (higher API gravity), low sulfur contents and high H/C ratios signify higher quality crude oil products, and thus yield higher product prices ^[5].

The construction duration for the SHC unit is approximated to be 5 years. A project lifetime 35 years will be used to determine the return on capital value. A 15% hurdle rate will be used as a benchmark to determine the feasibility of the project. The project will be evaluated in \$CAD and the location factor of 1.35 will be used. A list of project requirements for this project includes meeting the desired capacity, applicable safety regulations, and emission limits. A list of optional priorities for this project revolve around optimizing operating utility cost, optimizing hydrogen recycles, product desulfurization, and catalyst recovery.

4. JUSTIFICATION

4.1 Product Yield

A significant financial advantage proposed by utilizing SHC upgrading is the added benefit of increased product yield compared to carbon rejection technologies. Delayed coking, for example, have volumetric product yields of 82% on average. Conversely, the proposed SHC plant has a volumetric product yield of 125%. This means that the proposed design can produce as much as 43% more product for a given feed compared to the average upgrader in Alberta.^[34] The main reason for this is that our process undergoes combined hydrotreating and thermal cracking, where carbon-rejection methods do only thermal cracking. The functional difference is that carbon-rejection methods produce a carbon-residue waste (called coke), whereas SHC does not.

Table 1: Price of Athabasca bitumen and SHC final product.

Component	Price (CAD/bbl) ^[34]
Bitumen	\$50
Crude Oil	\$75

4.2 Product Composition

The quality of the product produced in the SHC is the key economic driver for this project. Figure 2 shows a comparison graph of hydrocarbon product composition based on boiling fractions for Athabasca bitumen feed and the proposed synthetic crude product. From this graph we can see the significant shift in product composition from heavier products to lighter ones, specifically in the 524+ °C (vacuum residue) range. Conversion in this process is defined by the reduction in vacuum residue in the product:

$$\text{Conversion} = 1 - \frac{\text{mass of residue in product}}{\text{mass of residue in feed}}$$

Using this formula, the proposed process has a calculated conversion of 95%, and drops residue content from 50 wt.% in the feed to 2.5 wt.% in the reactor liquid. With the addition of the product separation processes, the final saleable product has a residue content of 1.1 wt.%. Additionally, the proposed product meets pipeline specification, so it can be directly sold as is without requiring further processing or diluting.

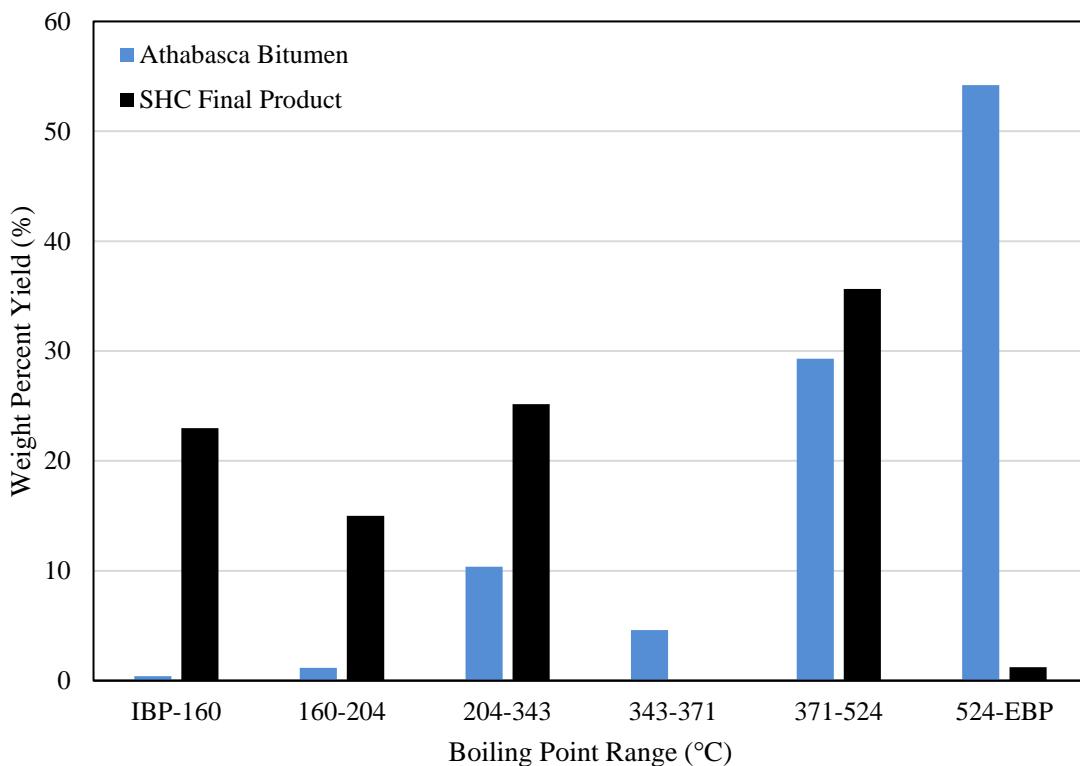


Figure 2: boiling fraction distributions for Athabasca bitumen and SHC final product.

5. ALTERNATIVES CONSIDERED

Based on the explicit nature of the project, the technologies evaluated will be different process options inherent in the SHC unit. The following sections will discuss technologies regarding catalyst choice, H₂S removal, and hydrogen production.

5.1 Do Nothing

The “do nothing” case pertains to no upgrading plant being built to process the available bitumen. Instead, the 100,000 bbl/d supply will be sent to an existing upgrading facility based on carbon-rejection technologies. This option acts as a basis of comparison for comparing SHC technology with existing technologies, but is outside the scope of our analysis, so it will not be considered further.

5.2 Homogeneously Dispersed Catalysts

Catalysts in this category are either water-soluble or oil-soluble metal compounds selected from elements of group IV through B–VIII. Oil-soluble catalysts are created using sulphides, or oxides (or combinations of both) or a salt of a group IV through group VIII metal (Mo, Ni, Co, W, Cr, V, Fe, Cu, Zn, etc.) along with organic acid ^[14]. However, the most commonly used metals are molybdenum, nickel, cobalt, and chromium, with Mo being the most catalytically active ^[10,14]. Industrial examples of this catalyst are iron pentacarbonyl, molybdenum 2-ethyl hexanoate (Alberta Research Co. (HC)₃ technology), and ammonium salts of vanadium sulfide (UOP Company) ^[7,8].

Advantages ^[8]:

- These catalysts achieve high conversion for heavy oils (90 wt.%).
- High catalytic activity allows for high levels of desulfurization to produce sweet products.

Disadvantages ^[8]:

- Catalysts of this type are highly expensive, meaning catalyst recovery is paramount in processing.
- Catalyst deactivation (and coke agglomeration) require catalyst regeneration steps, thus making the process more intensive.

5.3 Partial Oxidation (POX) of Heavy Hydrocarbons

The partial oxidation process uses oxygen to convert heavy hydrocarbons into a syngas mixture of hydrogen, carbon monoxide and carbon dioxide. The energy supplied to drive the process is sourced from firing the heavy feedstock [11]. This process requires a front-end gas separator to obtain high purity oxygen as reactant from air which is sent to the POX reactor and later purified in a pressure swing adsorber, similar to SMR process. This process is less efficient in comparison to the SMR process with a reaction efficiency of 50% [11]. The additional capital cost required due to the front-end gas separator makes the POX alternative less favourable for hydrogen production for the slurry hydrocracker process.

Advantages [11]:

- High hydrogen purity product (96-98wt.%).
- Can process crude product from cracked bitumen as reactant or fuel.

Disadvantages [11]:

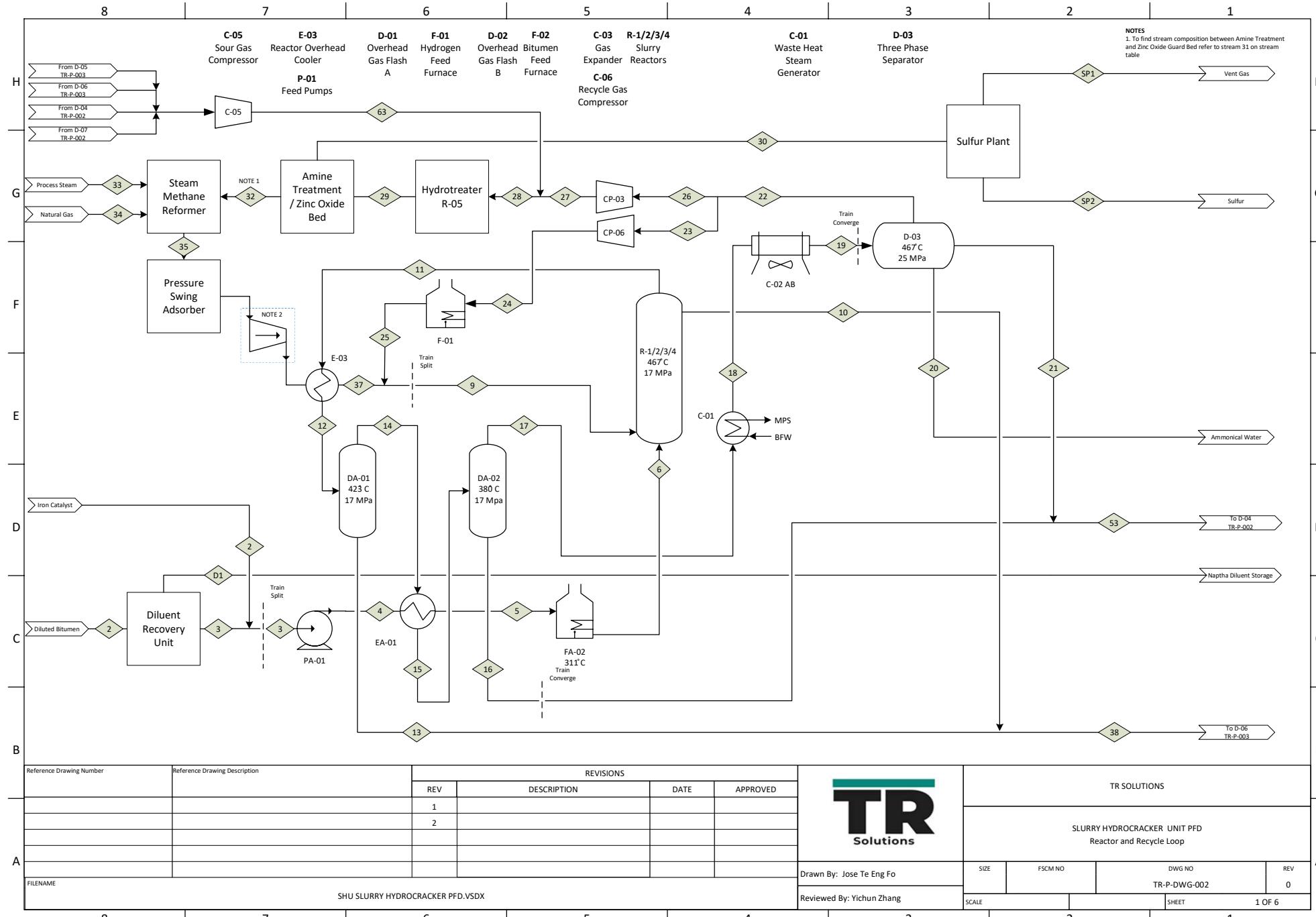
- High CO₂ emissions.
- Low efficiency.
- Higher Capital Cost due to front end air separation unit.

6. PROJECT DESCRIPTION

6.1 Process Flow Diagram

The process flow diagram (PFD) for the slurry hydrocracker unit is given in Figure 3. Table 1 is a summary stream table for all process lines. A detailed stream table with component mass fractions is located in Appendix D. Due to high flow rates, a large part of the proposed design is split into two trains (A & B), and the train convergence and splits are denoted in the PFD. Each train's equipment is denoted with the train letter after the equipment letter, e.g. EA-01 signifies heat exchanger 1 on train A.

Figure 3: Process flow diagram of proposed design.



Reference Drawing Number	Reference Drawing Description	REVISIONS			
		REV	DESCRIPTION	DATE	APPROVED
		1			
		2			
A					
FILENAME					

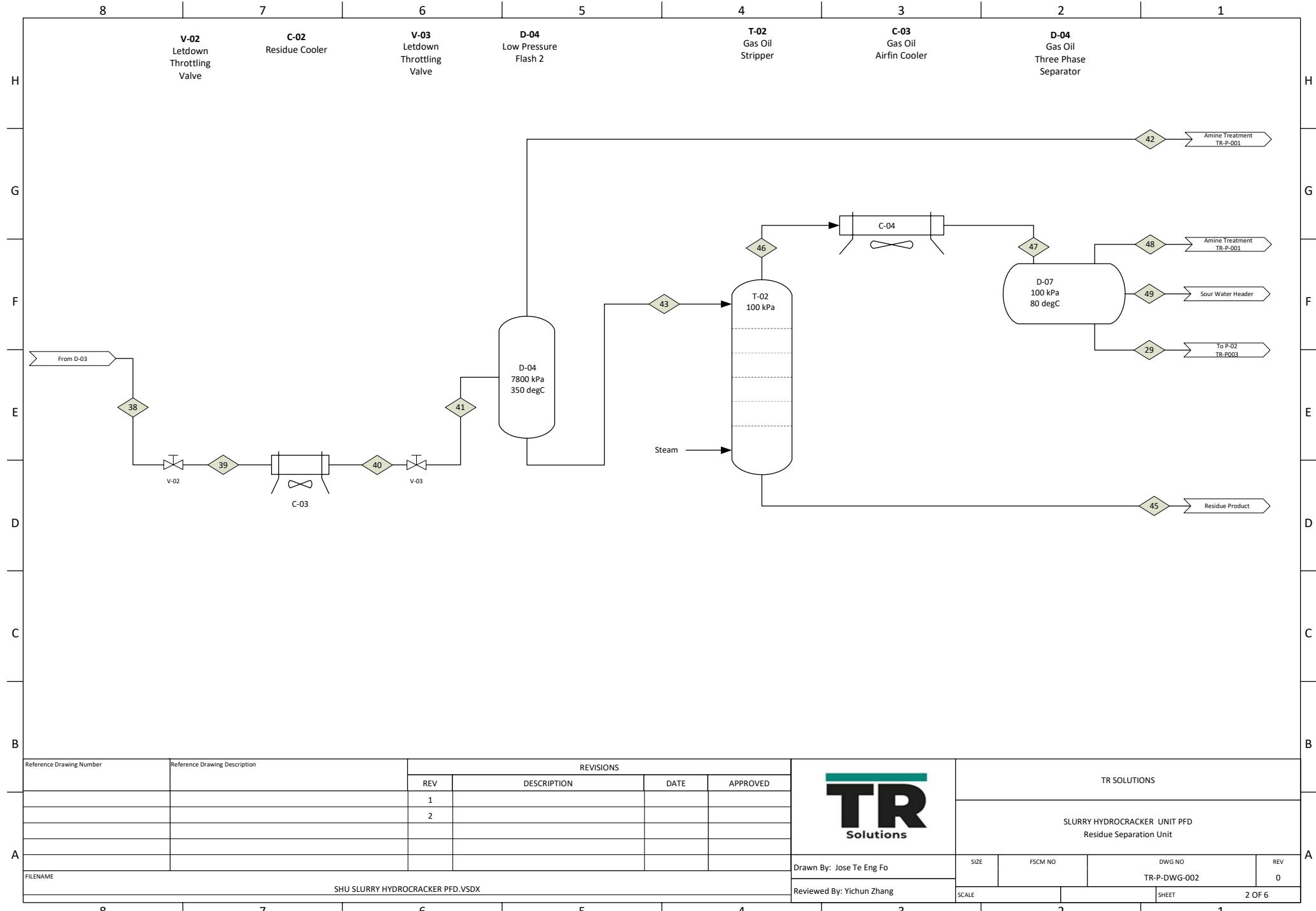


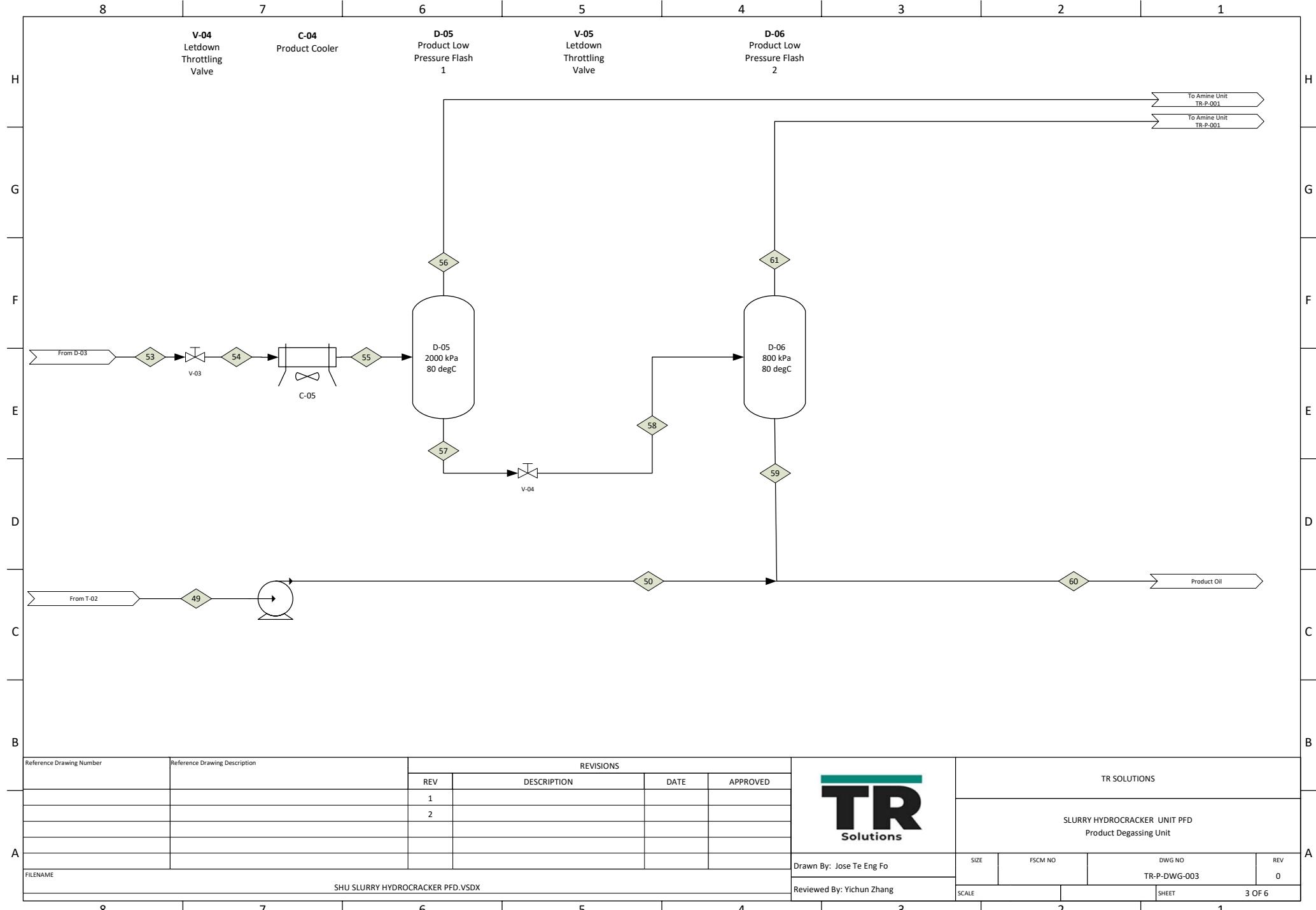
TR SOLUTIONS

SLURRY HYDROCRACKER UNIT PFD

Reviewed By: Yichun Zhang

Drawn By: Jose Te Eng Fo	SIZE	FCSM NO	DWG NO TR-P-DWG-002	REV 0
Reviewed By: Yichun Zhang	SCALE		SHEET	1 OF 6





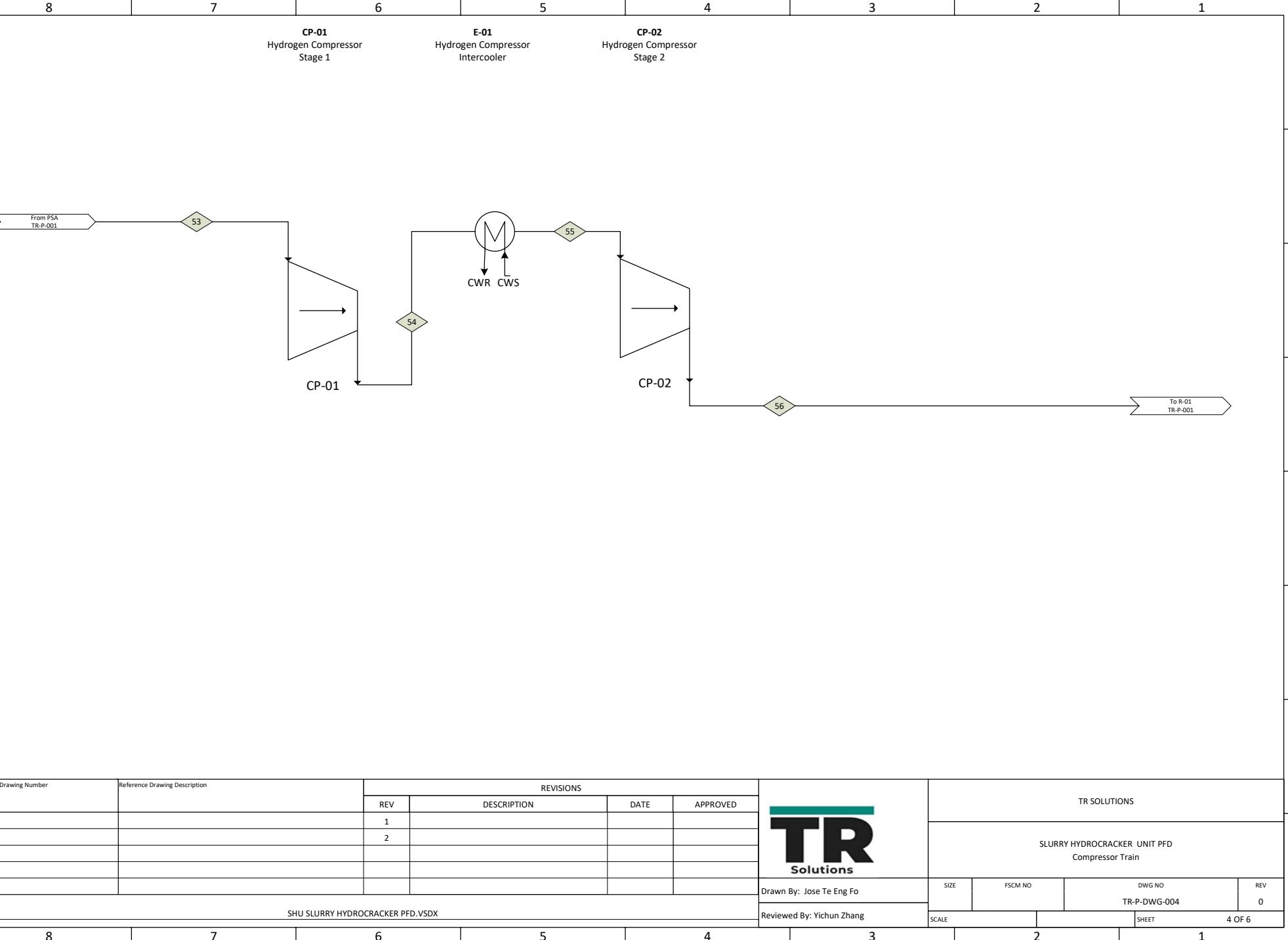


Table 2: Summary stream table.

Stream Property	S1	S2	S3	S4	S5	S6	S7	S8	S9	S10	S11	S12	S13	S14	S15
T (°C)	79.3	80.0	80.0	85.5	244.3	360.0	360.0	360.0	354.8	466.9	466.9	450.0	450.0	450.0	380.0
P (kPa)	500.00	500.00	500.00	17200.0	17100.0	17000.0	17000.0	17000.0	17000.0	17000.0	17000.0	16900.0	16900.0	16900.0	16800.0
Vapour Fraction (-)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.00	1.00	1.00	0.9990	0.00	1.00	0.9660
Mole Flow (k·mol/h)	6523.72	2320.85	4033.40	4033.40	4033.40	4033.40	43.81	4077.21	38643.0	50.37	31985.5	31985.5	30.76	31954.8	31954.8
Mass Flow (kg/h)	865516	200000	665516	665516	665516	665516	6655.16	672171.	180229	9175.69	843224	843224	10164.1	833060	833060
Mass Density (kg/m³)	-	-	892.193	895.211	718.377	606.270	-	-	14.5758	647.153	68.8459	70.1394	622.690	69.3881	76.9617
Volume Flow (m³/h)	946.351	200.419	745.932	743.417	926.415	1097.72	-	-	12364.9	3.895	12247.9	12022.1	16.323	12005.7	10824.3

Stream Property	S16	S17	S18	S19	S20	S21	S22	S23	S24	S25	S26	S27	S28	S29	S30
T (°C)	380.0	380.0	240.0	70.0	70.0	70.0	70.0	70.0	72.7	366.0	70.0	10.1	119.0	50.0	75.0
P (kPa)	16800.0	16800.0	16750.0	16700.0	16700.0	16700.0	16700.0	16700.0	17100.0	17000.0	16700.0	5500.00	5500.00	5500.00	25000.0
Vapour Fraction (-)	0.00	1.00	0.9361	0.8660	0.00	0.00	1.00	1.00	1.00	1.00	1.00	0.9908	1.0000	0.9969	0.00
Mole Flow (k·mol/h)	1085.75	30869.0	30869.3	30869.0	144.83	3991.25	26732.9	18713.0	18713.0	18713.1	8019.88	8019.88	9620.80	9310.50	726.21
Mass Flow (kg/h)	187409	645650	645646	645650	2608.63	442967	200075	140052	140052	140052	600224	600224	93067.0	93067.0	22732.0
Mass Density (kg/m³)	504.131	61.7694	79.6282	121.058	938.118	665.989	42.8839	42.8839	43.4764	23.0071	42.8839	17.8198	16.3563	20.9503	669.085
Volume Flow (m³/h)	371.748	10452.6	8108.26	5333.40	2.781	665.127	4665.50	3265.85	3221.34	6087.35	1399.65	3368.30	5689.97	4442.28	33.975

Stream Property	S31	S32	S33	S34	S35	S36	S37	S38	S39	S40	S41	S42	S43	S44	S45
T (°C)	40.0	36.6	300.0	200.0	80.0	80.0	342.0	453.3	455.3	350.0	350.0	350.0	350.0	100.0	197.2
P (kPa)	5400.00	2600.00	2600.00	2600.00	500.00	2100.00	17000.0	16900.0	7900.00	7900.00	7800.00	7800.00	7800.00	100.00	100.00
Vapour Fraction (-)	1.0000	1.0000	1.00	1.00	1.0000	1.00	1.00	0.0011	0.1857	0.1867	0.1881	1.00	0.00	1.00	2.2E-08
Mole Flow (k·mol/h)	8584.29	8584.29	7715.67	2162.32	10551.8	19930.0	19930.0	81.13	81.13	81.13	81.13	7.02	74.11	249.79	53.56
Mass Flow (kg/h)	70335.0	70335.0	139000	39845.0	207918	40176.5	40176.5	19339.8	19339.8	19339.8	19339.8	126.88	19212.9	4500.00	14424.4
Mass Density (kg/m³)	17.1653	8.3304	10.5533	12.3105	3.3625	1.4330	6.4301	611.719	498.764	596.776	594.259	26.8425	755.648	0.5857	1052.94
Volume Flow (m³/h)	4097.50	8443.16	13171.2	3236.67	61835.3	28036.2	6248.22	20.736	25.432	21.255	21.345	4.727	16.619	7683.64	7.379

Stream Property	S46	S47	S48	S49	S50	S51	S52	S53	S54	S55	S56	S57	S58	S59	S60
T (°C)	319.0	80.0	80.0	80.0	80.1	80.0	75.9	172.8	173.0	80.0	80.0	80.0	78.2	78.2	78.2
P (kPa)	100.00	90.00	90.00	90.00	490.00	90.00	90.00	16700.0	2000.00	2000.00	2000.00	2000.00	400.00	400.00	400.00
Vapour Fraction (-)	1.0000	0.0310	1.00	0.00	0.00	0.00	0.0048	0.0018	0.3498	0.2481	1.00	0.00	0.0854	0.00	0.00
Mole Flow (k·mol/h)	270.34	270.34	8.38	16.59	16.59	245.37	390.19	5077.00	5077.00	5077.00	1259.58	3817.42	3817.42	3491.47	3508.07
Mass Flow (kg/h)	9288.51	9288.51	175.28	4692.76	4692.76	4420.47	7029.10	630377	630377	630377	20832.7	609544	609544	597634	602327
Mass Density (kg/m³)	0.6999	32.9052	0.6444	821.100	821.206	971.606	104.029	580.410	155.012	238.016	11.5326	723.881	194.752	735.753	736.343
Volume Flow (m³/h)	13272.1	282.281	272.016	5.715	5.714	4.550	67.569	1086.09	4066.63	2648.47	1806.42	842.050	3129.85	812.276	817.997

Stream Property	S61	S62	S63	CT-S01	CT-S02	CT-S03	CP-S01
T (°C)	78.2	76.0	460.0	245.3	80.0	250.0	100.0
P (kPa)	400.00	90.00	5500.00	5985.00	5885.00	17100.0	100.00
Vapour Fraction (-)	1.00	1.0000	1.00	1.00	1.00	1.00	1.00
Mole Flow (k·mol/h)	325.95	1600.93	1600.93	19930.0	19930.0	19930.0	1407.48
Mass Flow (kg/h)	11909.8	33044.6	33044.6	40176.5	40176.5	40176.5	23797.0
Mass Density (kg/m³)	5.1389	0.6412	18.5003	2.7551	3.9682	7.5660	-
Volume Flow (m³/h)	2317.57	51536.4	1786.17	14582.5	10124.5	5310.10	-

6.2 Process Detail

6.2.1 Bitumen Feed

Diluted bitumen feed enters the facility in the DRU at 143,000 bbl/day at 1 atm and 80°C. The DRU is expected to recover 100,000 bbl/day equivalent of hot bitumen feed for upgrading where iron (II) sulfate catalyst. The bitumen is raised to reactor conditions (17MPa; 467°C) through a multistage pump (P-01) and by preheating with a reactor effluent heat exchanger (E-01) and a fired heater (F-01). The bitumen is fed to the slurry bubble column reactors (R-01/02/03/04) with sparged hydrogen to conduct hydrocracking.

6.2.2 Hydrogen Loop

The slurry bubble column reactors vaporize the majority of the cracked product, meaning it leaves with the vapor overheads (alongside the unreacted hydrogen and other process gases). The hydrogen-rich vapour stream is cooled in three stages with high pressure flash separations between each stage. The first stage is cooled by the hydrogen feed in E-03, and condensate is separated through D-01. This is repeated by the bitumen preheat exchange in E-01 and D-02 respectively. The last stage of cooling is conducted by a waste-heat boiler and air-fin cooler combo (C-01) to 70°C. This stream contains a hydrogen-rich mix of non-condensable gases, liquid hydrocarbons and liquid water, so it is separated by a three-phase separator. The hydrogen rich gas will leave the overhead of the three-phase separator and is split into two streams. One stream is compressed by C-06 and heated by F-01 to reactor conditions, and acts as a hydrogen recycle to maintain high hydrogen partial pressure inside the reactors. The second stream is expanded by C-03 and is mixed with various off gasses from low pressure separators (D-05, D-06, D-07, D-08). The overall stream passes through nickel-based packed bed, R-05, to react the residual amounts of sulfur and nitrogen aromatic-containing compounds to produce H₂S and NH₃. This stream is then treated by an Amine unit, followed by a zinc oxide guard bed to completely remove hydrogen sulfide prior to sending the stream to the SMR. The SMR unit combines light hydrocarbons (methane to butane) with steam and reforms them to hydrogen gas. Natural gas make-up is also sent to the SMR to produce the required amount of hydrogen. Of the hydrocarbon streams entering the SMR, 50 mol% is make-up natural gas, and 50 mol% is reactor-off gas. The hydrogen stream from the SMR is purified using the PSA. While the PSA vent gas is burned as a utility heating stream, the purified hydrogen stream is compressed by the compressor train, CT-

C01 and CT-C02, and is preheated by E-03 before mixing with the internal recycle to be sparged into the reactors.

6.2.3 Product Separation

The desired light hydrocarbon product is collected in the three-phase separator D-03 and is removed through the hydrocarbon draw. The water draw of D-03 will have dissolved ammonium salts due to the acid base reaction of ammonia and hydrogen sulfide. The product stream is cooled by an airfin cooler, C-03, and is throttled to a lower pressure with throttling valves V-02 and V-03 in two stages with flash separations in between each stage (using D-04 and D-05). The non-condensable vapor is recovered in the overhead of each flash drum and is compressed to be fed into reactor R-05 (and meet with the vapour products from the three-phase separator). The pressures and temperatures of each separation unit were determined to achieve a Reid Vapor Pressure specification of 96 kPa in the product liquid oil.

6.2.4 Residue Purification

The slurry bubble column reactor has a liquid draw for the product stream that does not vaporize. This stream predominantly contains high density residue gas oil and spent catalyst. This stream is throttled to lower pressures and sent to flash drum D-06 to recover dissolved light ends and is then steam-stripped in T-02 to recover gas oil from the residue stream. The gas oil recovered is pumped to equivalent pressures and mixed with the product crude. The residue is sent to a nearby coking unit OSBL.

6.3 Process Chemistry

6.3.1 Hydroconversion (Bitumen Upgrading) Reaction

As previously mentioned, utilizing SHC is beneficial over conventional methods of carbon rejection because of the high rates of residue conversion, and high volumetric product yield. The reason for these attributes is due to the upgrading reaction(s) taking place. In carbon-rejection processes, upgrading is achieved mainly through thermal conversion (cracking) reactions, whereas SHC performs residue hydroconversion, which combines catalytic hydrogenation and thermal cracking [5].

In straight thermal cracking, large, heavy hydrocarbon molecules are decomposed (cracked apart) into smaller components through the addition of heat. Although this technology produces reasonably high-quality products, a significant disadvantage of this technology is the production

of low-quality carbon residue (waste) called coke. In short, coke is formed as a result of hydrogen disproportionation (HD), or hydrogen transfer, taking place between components [5]. As HD occurs, some products gain hydrogen and become hydrogen enriched, while others lose hydrogen and become hydrogen depleted. To this end, some molecules become so hydrogen depleted that they become insoluble and form solids. These solids are known as coke and are essentially a waste product. In practise, delayed coking, for example, can convert 25 wt.% of the bitumen feed into coke, signifying significant product losses [5].

Although our design undergoes thermal cracking, it prevents coke formation through the act of catalytic hydrogenation. Rather than relying on HD to enrich molecules with hydrogen, catalytic hydrogenation utilizes a catalyst and hydrogen gas to add hydrogen to products. In this, hydrogen and catalyst work together to hydrogenate carbon compounds and inhibit coke formation by hydrogenating the coke precursor and removing heteroatoms [5]. The combination of hydrogenation and heteroatom removal significantly increases product hydrogen to carbon ratio, and therefore product quality.

In addition, it was found that increasing levels of hydrogen partial pressure will decrease coke formation but will also decrease yields of gas and naphtha (while increasing yields of diesel and vacuum gas oil) [5]. According to Benham and Pruden (1996), CANMET hydrocracking processes typically have a hydrogen partial pressure of around 1500 psi (10.34 MPa) [18]. Hence, it was decided that this project will maintain a hydrogen partial pressure of 11 MPa.

6.3.2 Catalyst Use

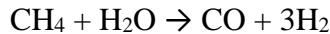
As previously mentioned, catalytic hydrogenation utilizes hydrogen and a catalyst is to saturate unsaturated carbon bonds and break aromatic compounds. Catalyst properties affect rates of coke formation, incompatible product formation, and impurity removal (sulfur, nitrogen and heavy metals) [7,8]. The potential catalysts for use are either dispersed heterogeneous (usually iron-based) or homogeneous (usually molybdenum-based). The catalyst of choice for this process is iron (II) sulfate, FeSO₄, a heterogeneous solid catalyst. It is characterized as a cheap, low activity catalyst. The main advantages of this catalyst are that it can achieve a 90% conversion of heavy oil, and that it is single-use and disposable (and therefore does not require catalyst separation or regeneration) [6]. The addition of iron sulfate also allows coke-free operation at a lower hydrogen partial pressure for given operating temperature and pressure [5]. The disadvantages of this catalyst are that low catalytic activity and low rates of dispersion provide low rates of desulfurization and

require large amounts of catalyst (1-5 w.t%) to be used. The consequences of high catalyst doses are heightened rates of equipment wear and product contamination. The catalyst will agglomerate with the heavy unconverted residue in the product which can be separated as bottoms of a tray column as a partial recycle to the reactor with a purge. This internal recycle option is presently demonstrated by an existing facility achieving the total conversion of feed [9]. This will be evaluated as a potential variable option within the hydrocracking process to optimize the overall process.

6.3.3 Steam Methane Reformer (Hydrogen Production)

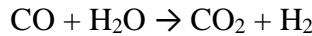
The steam methane reforming process uses natural gas (composed primarily of methane) to produce syngas that contains hydrogen gas and carbon monoxide [11].

The principal reaction of steam reforming of methane is:



However, larger hydrocarbons will undergo a similar reaction to produce the stoichiometric amount of carbon monoxide and hydrogen gas.

In addition to the steam-reforming reaction, additionally the water-gas shift reaction is performed to increase hydrogen yield. This reaction converts carbon monoxide and water to carbon dioxide and hydrogen gas:



Although both reactions are run in parallel, the dominant reaction is the hydrocarbon steam reforming, so the main products of these reactions are hydrogen and carbon monoxide, with some amounts of carbon dioxide and unreacted water remaining.

The streaming reforming reaction involved in the process is reversible and highly endothermic, therefore the generation of H₂ is favoured at high temperatures, high steam to methane ratios, and low pressures. Typical industrial conditions of the steam reformer are at 26 bar and have a feed steam to methane ratio of 2.5 [12].

6.4 Simulation

Specific details relating to simulation development, detailed design justifications, simulation flowsheets and simulation files are contained in Appendix B.

6.4.1 Reactor Component Balance

Due to complexity of modelling a sparged slurry bubble column reactor and limited software capabilities. An elemental mass balance was completed around the reactor, between Athabasca bitumen and the products being specified by CANMET yields. Detailed calculations are shown in Appendix C.

The reactor mass balance shows that with 19.8% of naphtha, 33.5% of LGO, 28.5% of VGO, 4.5% of vacuum residue, and 13.7% of gas product being produced from the bitumen feed, 27 tons of hydrogen gas will be consumed in the reaction ^{[5][6]}. This provides the basis for estimating the SMR duty. Additional hydrogen will pass through as an inert, such that a hydrogen partial pressure is maintained at 11 MPa to suppress coke formation. It is also predicted that this process will achieve 69% of desulfurization and 31% of denitrogenation.

6.4.2 Thermodynamic Model Selection

The thermodynamic model of choice for simulation development was the refinery characterization of the Advanced Peng Robinson (APR) equation of state. This decision was made with the Decision Tree Approach for phase behaviour from ChE 465 class notes. To characterize the bitumen feed and product crude, Symmetry was used to simulate the oils with PIONA (Paraffin, Iso Paraffins, Olefins, Naphthenes, Aromatics) analysis.

6.4.3 Bitumen and Product Oil Characterization

For the purposes of this project, the product oil simulation was simplified to characterizations only based on paraffins, nitrogen aromatics and sulfur aromatics. Boiling point cuts are dependent on paraffin chain length and aromatic chain length and its connected heteroatom. The aromatic slate component for each carbon chain length section is distributed linearly in each C_n cut to match the boiling point curve paralleled to CANMET ^[6] product composition as stated in Appendix B. Similarly, the bitumen was simulated to align with available Athabasca Bitumen boiling curves (reference). The aromatic characterization assumes a 50 wt.% sulfur and 50 wt.% nitrogen distribution for the corresponding aromatic chain. The paraffin boiling point cuts are simulated using correlations using correlations by built in correlations and are shown in Table 3 ^[19].

Table 3: Paraffin boiling cuts.

Cn Groups	Initial BP (°C)	Final BP (°C)	# of Cn Groups	Cn Distribution for each group (wt.%)	Total Distribution (wt.%)
C ₆ - C ₁₀	36.6	151.3	5	5.3	26.6
C ₁₁ - C ₁₄	151.3	254	4	4.9	19.6
C ₁₅ - C ₁₉	254	331.1	5	5.27	52.7
C ₂₀ - C ₃₅	331.1	490.6	15	0.084	0.92
C ₃₆₊	490.6	808.1	8	0.165	0.13

6.4.4 Model Limitations

Although great care was taken to ensure reasonably accurate output, there are known limitations to the simulation. In some cases, specific equipment is not modelled in the simulation, like the Claus (sulfur) plant and the zinc oxide bed, and bubble column. In other cases, equipment is only partially simulated, like the amine unit, SMR and PSA. In these cases, process detail is either adapted to reflect the real process (where no equipment exists), or equipment operating conditions are simulated based on literature research or generalized simulations (but is not specifically optimized). The justification for these limitations is that these process units fall outside the scope of detailed engineering work.

Another known limitation of the simulation is the limited oil characterization performed. The hydrocarbon stream's physical properties in the model are directly dependent on the simulated components and its composition. The model was simplified using key assumptions in the PIONA analysis to estimate characteristics of the bitumen and product oil due to lack of experimental data. These limitations can cause errors in estimating physical properties such as heat capacity, density, and compressibility ratio which can propagate to equipment sizing calculations. From the model validation, there is significant deviance between the bitumen modelling and the commercially known properties of Athabasca Bitumen. However, due to the separated nature of the modelling, this error is only significant in the inlet bitumen feed section of the design, including feed pumps, heat exchanger E-01 and furnace F-02.

The model validation conducted resulted in a difference of 10% between the specific gravities of literature and simulated Athabasca Bitumen streams. This is an acceptable deviation at this stage of design and analysis of the SHC.

6.5 Overall Mass and Energy Balances

The overall mass balance for the process is shown below in Table 4. Further details to support the mass balance, and additional information about the mass balance around the slurry bubble column reactors is located in Appendix A. The difference between total mass in and out is 0.24 kg/h, and is known to be around the reactor, due specifically to the mass of hydrogen produced in the SMR.

Table 4: Overall mass balance for the proposed design.

In		Out	
Stream Description	m (kg/h)	Stream Description	m (kg/h)
Bitumen Feed	665,516	Product Gas	91,369
(Fresh) Catalyst	6,655	Naphtha	128,937
Fresh H ₂	40,176	LGO	242,047
Recycle Gas	140,052	VGO	215,123
		Residue	14,716
		(Spent) Catalyst	6,655
		Unreacted H ₂	13,500
		Recycle Gas	140,052
Total:	852,400	Total:	852,399

The overall energy balance for the process is located below in Table 5. Further details to support the energy balance, and additional information about the energy balance around the slurry bubble column reactor is located in Appendix A. Stream enthalpies were taken directly from the simulation and appear positive or negative due to the arbitrary enthalpy baseline set by VMG.

Table 5: Overall energy balance for the proposed design.

Stream Description	Stream Number	Electricity Input (kW)	Heat In (kW)	Heat Out (kW)	Energy In (kW)	Energy Out (kW)
Bitumen Feed	S3	-	-	-	-11,814	
Steam	S33	-	-	-	39,592	
Methane	S34	-	-	-	10,800	
Stripper Steam	S44	-	-	-	862	
Final Product	S60	-	-	-		3,165
Residue Product	S45	-	-	-		430
SMR Liquid	SMR-S09	-	-	-		506
Acid Gas	S30	-	-	-		712
PSA Purge	S35	-	-	-		30,559
Waste-Water	S52	-	-	-		3,282
Equipment Description	Equipment Number	Electricity Input (kW)	Heat In (kW)	Heat Out (kW)	Energy In (kW)	Energy Out (kW)
Slurry Bubble Column Reactor(s)	R-01 – R-04	-	93,536	-	93,536	
Conversion Reactor	R-05	-	-	8,151		8,151
Pump Feed Train	P-01	4,965	-	-	4,965	
Final Product Pump	P-02	1	-	-		1
Compressor Train Compressors	CT-CP01 / CT-CP02	54,734	-	-	54,734	
Compressor Train Cooler	CT-C01	-	-	26,753		26,753
Product Gas Expander	EX-01	-	-	-		4,642
Product Overheads Cooler	C-01	-	-	211,564		211,564
Product Bottoms Cooler	C-02	-	-	1,030		1,030
Stripper Cooler	C-03	-	-	4,559		4,559
3-Phase Separator Cooler	C-04	-	-	41,843		41,843
Bitumen Feed Furnace	F-01	-	57,014	-	57,014	
Internal Recycle Gas Furnace	F-02	-	56,431	-	56,431	
Bitumen Feed Pre-heater	E-01	-	64,949	-		
Hydrogen Pre-Heater	E-02	-	15,038	-		
Internal Recycle Gas Compressor	CP-01	480	-	-	480	
Stripping Unit Overheads Compressor	CP-02	9,581	-	-	9,581	
SMR/PSA	-	-	-	-	12,924	
Amine Unit	AU	-	-	-		3,115
					Total:	329,105
					Difference:	-2,204
						1%

6.6 Equipment Selection

A detailed summary of equipment specifications and operating parameters is included in Appendix C.

6.6.1 Equipment List

The table below summarizes the equipment list for the SHC process. It shows the equipment number from PFD, equipment description, material of construction and size and/or capacity specifications.

Table 6: Equipment List with equipment descriptions.

Equipment Number	Equipment Description	Equipment Sub-Type	Material	Capacity / Size / Duty Specifications	Total Number of Equipment
C-01	Waste Heat Boiler	Floating Head	CS/SS	A = 924.5 m ²	8
C-02	Cooler	Air fin	Stainless Steel	A = 1827 m ²	2
C-03	Cooler	Air fin	Stainless Steel	A = 84 m ²	1
C-04	Cooler	Air fin	Stainless Steel	A = 1638 m ²	1
CP-01	Compressor	Centrifugal	Stainless Steel	w _s = 480 kW	1
CP-02	Compressor	Centrifugal	Stainless Steel	w _s = 9,581 kW	1
CT-C01	Cooler	Air fin	Carbon Steel	A=161.4 m ²	1
CT-CP01	Compressor	Centrifugal	Stainless Steel	w _s = 26,777 kW	1
CT-CP02	Compressor	Centrifugal	Stainless Steel	w _s = 27,957 kW	1
D-01	Flash Drum	Vertical	Stainless Clad	D = 4 m L = 20 m	4
D-02	Flash Drum	Vertical	Stainless Clad	D = 4 m L = 20 m	4
D-03	3 Phase Separator	Horizontal	Stainless Clad	D = 3 m L = 7.08 m	1
D-04	Flash Drum	Vertical	Stainless Clad	D = 1.5 m L = 7.5 m	1
D-05	Flash Drum	Vertical	Stainless Clad	D = 4 m L = 20 m	4
D-06	Flash Drum	Vertical	Stainless Clad	D = 4 m L = 20 m	2
D-07	3 Phase Separator	Horizontal	Stainless Clad	D = 1 m L = 1.1 m	1
E-01	Heat Exchanger	Floating Head	CS/SS	A = 895 m ²	2
E-02	Heat Exchanger	U Tube	CS/SS	A = 30 m ²	2
EX-01	Gas Expander	Axial	Stainless Steel	ws = 2640 kW	1
F-01	Heater	Fired Heater		Q = 31627 kW	2
F-02	Heater	Fired Heater		Q = 56431 kW	2
P-01	Pump	Centrifugal	Cast Steel	ws = 769 kW Q = 0.104 m ³ /s	2
P-02	Pump	Centrifugal	Cast Steel	ws = 0.85 kW Q = 0.0016 m ³ /s	1
R-01	Main Reactor	Vertical	Stainless Steel	D = 4 m L = 30.8 m	8
R-05, R-06	Conversion Reactor	Vertical	Stainless Steel	D = 4 m L = 19.7 m	1
T-02	Stripper Tray Column		Stainless Steel	D = 1.5 m L = 4.04 m	1

6.6.2 Reactors

The design details, assumptions, and sample calculations for the Slurry Bubble Reactor can be found in Appendix E.

The slurry bubble column consists of a multiphase system in which gas and liquid phases are contacted with a solid phase (catalyst) promoting hydrogenation of the thermally cracked hydrocarbons. For this proposal it requires eight identical reactors in shape and design, with four reactors in parallel on two trains alongside. The size of the bitumen conversion reactors was based on residence time of 35 minutes. The volume of the entire column is 387 m³, the internal column diameter is 4 m, the tangent height 31 m, and the wall thickness 0.828 m. The material selected for the construction of each reach is driven by the process temperature and strength due to corrosion, therefore the material used is stainless steel.

6.6.3 Heat Exchangers

Heat exchangers were placed to recover heat from reactor effluent to reactor feeds. The complete list of all heat exchangers with their corresponding service and material are listed in the equipment list in Appendix E. The material of construction selected for each heat exchanger is stainless steel (SS) driven by the high amounts of hydrogen sulfide in the process streams. Sulfidic corrosion is a large problem especially at high temperatures ^[16].

Shell and Tube Exchangers (TEMA)

There are two shell and tube heat exchangers in the proposed design. The first preheat exchanger E-01 is a U-tube heat exchanger due to its relatively small area requirement and relatively clean shell and tube service. The U bend will allow for thermal expansion to relieve the thermal shock from changing reactor conditions. The E-02 is a fixed tube floating heat exchanger. The bitumen feed can cause excessive fouling and will therefore be flowing in the tube side for ease of cleaning. Expansion joints and the floating head are installed to relieve thermal expansion and stress. The waste heat steam generator C-01 was designed as a floating head heat exchanger with a supplementary kettle as the steam drum. The overall heat transfer coefficients chosen for the above heat exchangers are obtained from Table 4.15a in Ulrich and Vasudevan (2004).

Air Cooled Heat Exchangers (ACHE)

ACHE configurations were chosen as the main cooling utility for the unit due to its lower operating expenses. There is minimal airside fouling and the tubes in each bank are easy to clean ^[16]. Coefficients for the Air-Cooled Heat Exchangers (ACHE) are obtained in Table 4.15b in Ulrich

and Vasudevan. A detailed design of an air-cooled heat exchanger was conducted as per Hudson products guidelines [32] shown in Appendix E.

6.6.4 Three-phase Separator

There are two three-phase separators in this design. Both used for the separation of light hydrocarbons gas phase, water, and oil liquid phases. Water is removed from the oil phase, and the oil is joining back to the product stream as part of the crude oil product. It was assumed that the three-phase separators are perfect columns in shape, and a five minutes retention time of the liquid phase was assumed. The diameter of the separator was assumed to be 3 meters first and then the length was calculated accordingly base on a five minutes retention time.

6.6.5 Pumps

This design proposal requires a total of three pumps, two of which are the bitumen feed pumps, P-01, and the last is pump P-02 used downstream of the three-phase separator D-07. Detailed sizing and costing are shown in Appendices E and I respectively. Centrifugal-type pumps were selected due to their operability at the specified temperatures and pressures [16].

The feed of bitumen is split into two trains, where each train is equipped with a separate feed pump, such that in case of train maintenance, the alternate train can continue to operate normally. Each feed pump is required to operate at a capacity of 0.104 m³/s with the suction and discharge pressures being 5 bars and 172 bars respectively. Due to the large pressure difference, the work needs to be distributed to three stages, each achieving a third of the pressure difference. Based on the assumption that every stage of the pump has an efficiency of 75%, the shaft work required at each stage are 137 kW, 446 kW, and 1452 kW respectively.

Pump P-02 is required to operate at a capacity of 0.0016 m³/s. At an efficiency of 75%, the shaft work required for this pump is 0.85 kW.

6.6.6 Compressors

A total of four compressors are used in the process, which are compressors CT-CP01, CT-CP02, CP-01, and CP-02 respectively. Detailed sizing and costing are shown in Appendices E and I.

CT-CP01 and CT-CP02 are connected in series in the compressor train downstream of the PSA unit. The work required by each of the compressors are 26777 kW and 27957 kW respectively, as being determined by VMG simulations. CP-01 and CP-02 are used to compress the gas leaving at the top of the separators, each being required to achieve a work of 480 kW and 9581 kW

respectively. Centrifugal type compressors were selected by the same approach as to specify pump types [16].

6.6.7 Piping

Detailed line sizing was performed for main process lines starting after the DRU and leading into the internal gas recycle loop. A detailed line designation table with sizing sample calculations can be found in Appendix E.

Table 7: Line designation table.

Stream Number	Nominal Pipe Diameter (in)	Pipe Schedule	Pressure Drop (kPa/100m)	Pipe Material	Material Grade
S3	8	20	35	Carbon Steel	B
S4	8	100	55	Carbon Steel	B
S5	8	100	68	Carbon Steel	B
S6	8	100	83	Carbon Steel	B
S7	4	160	n/a	Carbon Steel	B
S8	8	160	84	Carbon Steel	B
S9	14	120	19	Stainless Steel	TP304
S10	4	140	n/a	Stainless Steel	TP304
S11	18	140	26	Stainless Steel	TP304
S12	18	120	29	Stainless Steel	TP304
S13	1.5	80	51	Stainless Steel	TP304
S14	18	140	25	Stainless Steel	TP304
S15	18	120	32	Stainless Steel	TP304
S16	8	120	5	Stainless Steel	TP304
S17	16	120	27	Stainless Steel	TP304
S18	16	120	34	Stainless Steel	TP304
S19	18	100	39	Stainless Steel	TP304
S20	1	40	71	Stainless Steel	TP304
S21	8	100	84	Stainless Steel	TP304
S22	12	100	44	Stainless Steel	TP304
S23	12	100	21	Stainless Steel	TP304
S24	10	100	13	Stainless Steel	TP304
S25	10	120	28	Stainless Steel	TP304

6.6.8 Tray Columns

This theoretical number of trays was obtained as per the procedures outlined in (Wankat, 2007) [32] for a stripper column design. The gas oil was assumed as the solute and the steam as the vapor phase streams. Sizing of the tray column was conducted as per Ulrich and Vasudevan guidelines (2004) [16]. The sizing calculations are shown in Appendix E.

6.6.9 Control Valves

Three level control valves were sized as per the methods outlined by (Cornell, 2012)^[30]. For simplified sizing, the valve was chosen such that the operating flow Cv was in 20-30% valve travel. The sizing calculations are shown in Appendix E.

6.6.10 Pressure Safety Valve

The maximum allowable pressure for D-03 was calculated as per ASME Standard. The D-03 pressure relief valve PSV-001 was sized as per the methods outlined by (Hellemans, 2009)^[31]. The detailed sizing calculations are shown in Appendix E.

6.6.11 Fired Heaters

The fired heater duties were calculated using Symmetry. The duties were used to size the heaters accordingly as per U&V The sizing calculations are shown in Appendix E^[16].

6.7 Material Selection

Material selection for process vessels and piping is split between carbon steel and stainless steel (and in some special cases, cast steel). The desired standard material of choice is carbon steel based on its strength and relatively cheap cost, however due to the presence of H₂S and NH₃ in our process stainless steel is required for some equipment for corrosion resistance. Vessels and process pipelines in contact with H₂S-containing vapour were priced as either straight stainless steel, or as stainless steel clad (stainless steel lined carbon steel).

6.8 Utilities

Table 8 and 9 shows the utility requirements for the overall unit. The values were calculated as a combination of AACE Class V and Class IV estimates and can be found in Appendix I. The vent gas from the pressure swing adsorber which contains high amounts of CO and H₂ will be burned as heating gas with a calculated heating value. The manufacturing expenses entailed with utilities below are estimated as per the correlations in Ulrich and Vasudevan (2004)^[16].

Table 8: Equipment utility summary.

Equipment	Tag	Utility Type	Requirement	Units
Fired Heater	F-A01/F-B01	Natural Gas	1,618,191	GJ
Fired Heater	F-A02/F-B02	Natural Gas	1,779,596	GJ
Cooler	C-A01/C-B01	Boiling Feed Water	1,538,304	m ³
Cooler	C-A02/C-B02	Electricity	4,320,367	kWh
Cooler	C-02	Electricity	98,900	kWh
Cooler	C-03	Electricity	1,936,532	kWh
Cooler	C-04	Electricity	2,121,793	kWh
Pump	P-01	Electricity	4806	kWh
Pump	P-02	Electricity	1	kWh
Compressor	CP-A01/CP-B01	Electricity	3,784,500	kWh
Compressor	CP-02	Electricity	75,533,826	kWh
Compressor	CT-C01	Electricity	217,640,032	kWh
Compressor	CT-C02	Electricity	227,232,044	kWh

The Class V estimated utility cost utilized by the steam methane reformer, amine unit, Claus plant, and diluent recovery unit are shown in Table 8 using cost correlations outlined by Maples (2000) [25].

Table 9: Utility requirements for Class IV estimated equipment.

Unit	Electricity kWh	Fuel (GJ)	Steam (kg)	Cooling Water (m ³)
Amine Sweetening Plant	940,029	-	710,649,413	-
Hydrogen Plant	100,33	30,698,890	-	-
Sulfur Plant	6,151,798	-	5,232,012	2,910,885
DRU	23,46	4,951,227	532,160,614	-

The vent gas from the SMR is used as a utility heating stream due to its combustible content (CO and H₂) content. The heating value of this stream was calculated by quantifying the mol of CO and H₂ combustion to CO₂ and H₂O respectively and multiplying by the heat of reaction

released. Table 10 shows the utility balances of the available vent gas and natural gas make up. Table 10 shows the heating utility sources and the available energy.

Table 10: Heating utility sources summary.

Utility Source	Energy (GJ)	% of Total Heat Utility
PSA Vent Gas	19,588,641	55%
Natural Gas	16,061,475	45%

6.9 Project Execution Strategy

TR Solutions estimates the SHC project plant to have five development phases including: basis of the design, engineering phase, procurement phase, construction phase and commissioning phase are shown in the form of Gantt chart in Figure 5. The project duration is 5 years from the preliminary engineering details May 1st, 2020 to the project start-up June 1st, 2025. It is important to acknowledge that these dates are subjected to change due to unforeseen changes, mainly from the technical challenges and construction period. Further details for the project schedule strategy are provided in Appendix J.

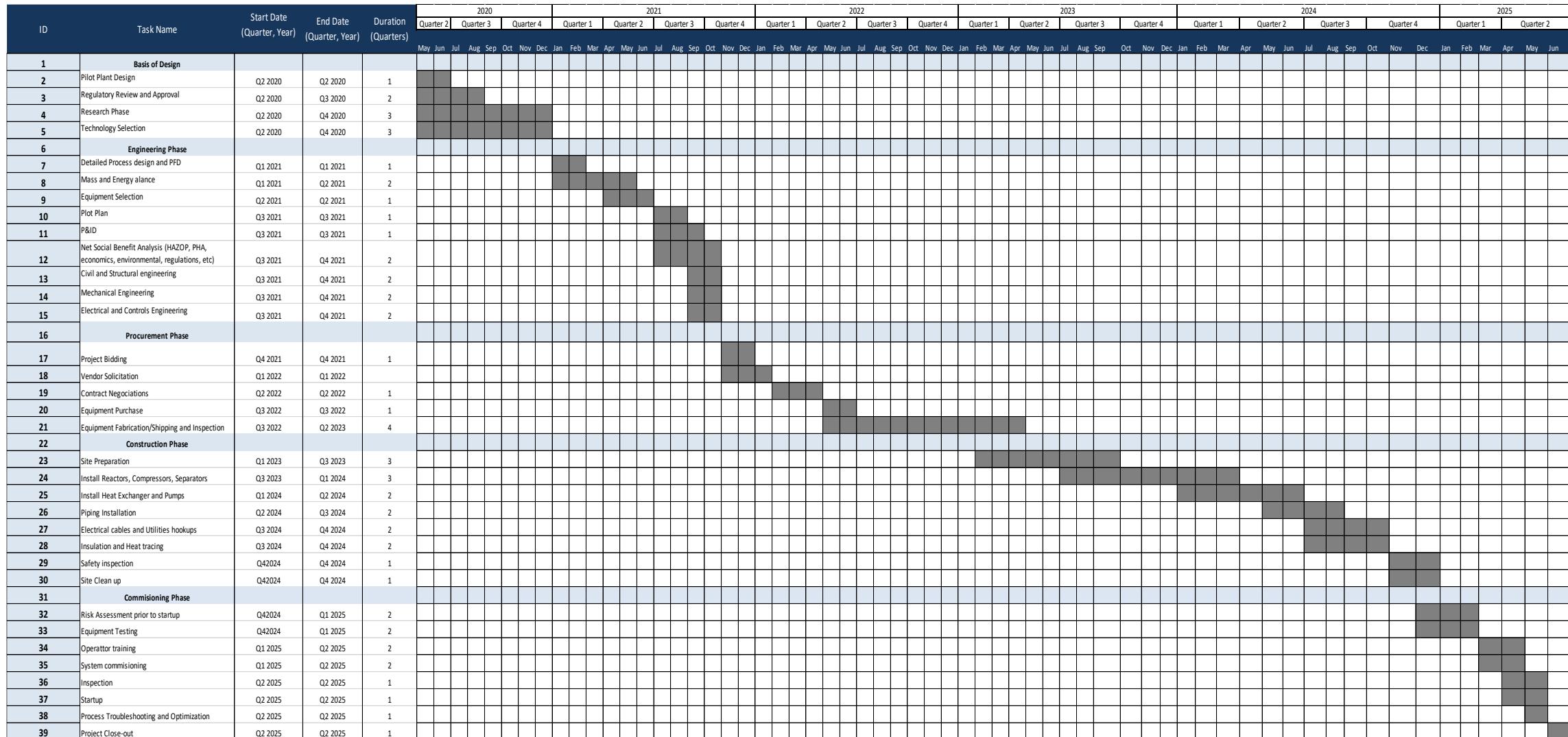


Figure 4: Process execution Gantt chart.

6.9.1 Engineering Plan

The engineering plan takes place after the conclusion of the design basis, research and selection of the technology. Upon approval of the design basic by Worley Limited the engineering phase of the project will start in January 2021 and is projected to take 10 months to complete. The following months will consist of in-depth detailed engineering including mass and energy balances, mechanical and electrical instrumentation PFDs, environmental and risk assessment, as well as the plot plan P&IDs. After these main stages, the engineering phase will continue alongside with the construction until the completion of the project.

6.9.2 Procurement Plan

Once the engineering drawing for the plant and economic analysis are completed the procurement phase will begin November 2021. During this time the project bidding, vendor solicitations, contract negotiations, equipment selection and purchase will be delivered in a timely manner. Equipment vendors will be first selected based on WorleyParsons past partnership, followed by local vendors which are preferred to due large equipment sizing and prevent unnecessary delays to the construction and installation phase. A period of approximately 12 months is allocated for the equipment and material manufacturing and shipment, and that takes the majority of the time allocated for the procurement plan.

6.9.3 Construction Plan

The construction phase begins in January 2023 and it is projected to be completed in January 2025, a total of 2 years. The early stage of the construction phase starts with below the ground foundation, pipeline layout, water and electricity installation, and site preparation while the manufacturing and shipping takes place as well. As the foundation is finalized and major equipment is shipped by July 2023 the second stage of the construction begins with the installation of reactor, compressors, separators and heat exchangers based on the plot plan developed during the engineering phase. Then the mechanical construction includes wiring, piping, and instrumentation is built around the major equipment. By the end of the year 2024 a safety inspection is performed throughout the entire plant to ensure that the project is finalized and prepared for the next phase.

6.9.4 Commissioning Plan

The commissioning phase is estimated to be 7 months from December 2024 to June 2025. This phase includes equipment testing, operator training, system commissioning, final inspections,

process troubleshooting and optimization. Once all tests and inspections are completed and meets the construction and engineering specifications the contractor will pass over the plant to TR Solution for the project close out.

7. ECONOMICS ANALYSIS

7.1 Overview

A grass-roots economic analysis for a 5-year construction period, followed by a 35-year period of operation was performed. This included a capital cost estimation, operating cost estimation, discounted cash flow and sensitivity analysis. Parameters and factors used for the cost calculations are listed in Table 11. Appendix I contains all economic analysis details, assumptions and sample calculations.

Table 11: Parameters and factors used for economic calculations.

Factors	Value
Location Factor (Edmonton)	1.35
Currency (CAD/USD)	1.34
CE Plant Cost Index in 2020	620
Operating Factor	0.9

7.2 Project Scope and Class Five Estimations

The economic scope of the project can be split to two sections: the hydrocracker unit and the auxiliary facilities. AACE Class IV Estimates were conducted on equipment that pertain to the hydrocracking unit. AACE Class V cost estimations were conducted on the SMR unit, the DRU, the amine unit and the Claus (sulfur) plant. Capital and operating costs for the Class V units were estimated (and adjusted to scale with project specifications) from literature values. Both capital costs and operating costs are adjusted from the example year to the year of 2020.

7.3 Capital Cost

Capital cost calculations including equipment costs can be found in Appendix I. For this project the total grass-root capital cost was estimated based on a project contingency of 20%, process contingency of 40% and a DFL adjustment of 70% of the bare module cost of the

equipment in order to allow a conservative estimation of project costs. Total grass-root capital cost was determined to be \$6.9 billion CAD. Equipment costs are estimated using methods in Ulrich and Vasudevan (2004) to obtain a U.S. Coast price in 2004. This price is then multiplied by Edmonton's location factor of 1.35 and CAD to USD ratio of 1.34 to obtain the actual price for each equipment^[16]. Table 12 summarises the cost for each type of equipment for the hydrocracker unit, this table does not contain information on class five cost estimations.

Table 12: Hydrocracking Unit Equipment cost summary.

Equipment Name	Number of Equipment	Total Cost
Reactors	8	\$125,762,081
Separators	17	\$114,687,434
Pumps	3	\$822,230
Compressors	4	\$253,742,921
Coolers	3	\$1,445,807
Fire Heaters	4	\$37,685,088
Heat Exchangers	4	\$1,372,814
Waste Heat Boilers	8	\$5,551,821
Expander	1	\$3,227,784
Air Fin Coolers	5	\$6,387,398
Total		\$551 M

The sum of module cost of all equipment draws a \$551 million CAD of total module cost for the hydrocracker unit. After including the class five capital cost estimations, a total module cost summary is shown in Table 13, the total of the SHC plant was found to be \$1.5 billion CAD.

Table 13: Class five capital cost summary.

Equipment Name	Cost	% of Total Cost
Hydrocracker Unit	\$551 M	37.6
Amine Treatment Unit	\$92 M	6.3
Hydrogen Plant (SMR)	\$589 M	40.2
Sulfur Plant	\$82 M	5.6
Diluent Recovery Unit	\$147 M	10.1
Zinc Oxide Bed	\$4 M	0.3
Total	\$1465 M	

This illustration on the breakdown of total module cost. It is evident that the SMR hydrogen plant incurs very high capital cost comparable to the overall hydrocracker unit, which makes the hydrogen production a barrier to the economic attractiveness of constructing a standalone SHC unit.

7.4 Operating Cost

It is assumed that all the costs and prices related to operation stay constant at January 2020 pricing throughout the 35-year operation. Only an inflation rate of 3% each year is considered. Prices for natural gas, bitumen, naphtha, crude oil product, sulphur product and operating labour wages are estimated based on 2019 average values. Unit prices for utilities are calculated based on methods from Ulrich and Vasudevan (2004) [16]. The unit price for each utility is then multiplied by its annual consumption to obtain the annual cost in 2020. Operating labour cost assumes an average hourly wage of \$33.25 CAD/hr with an operating factor of 0.9. The pricing for natural gas was obtained from the averaged Energy Alberta 2019 prices. A breakdown of operating cost is shown in Figure 5.

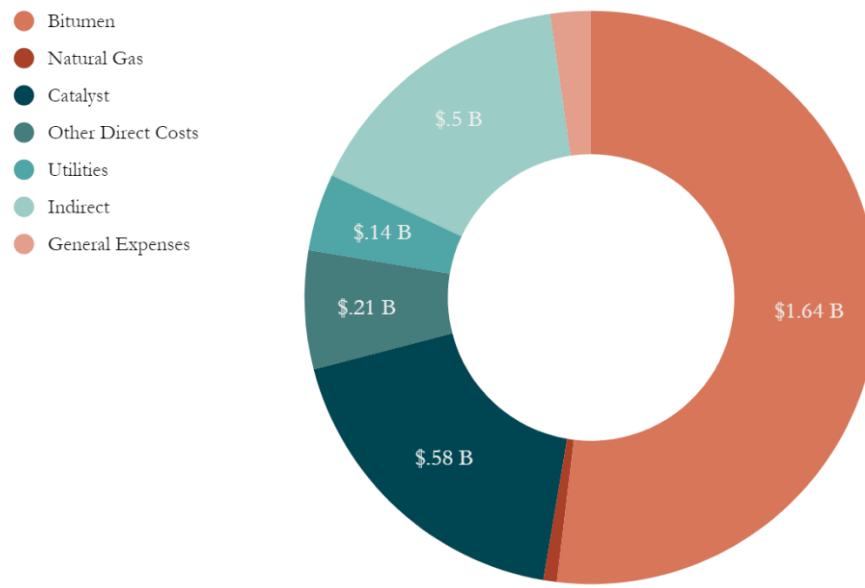


Figure 5: Annual operating cost breakdown.

As seen in the above Figure, bitumen cost takes up 68% of the total operating cost. The cost of 1.64 billion does not include the diluent purchased. Bitumen is priced at \$325.83 CAD/m³ from the 2019 statistics data [27].

7.5 Revenue

The revenue from this project consists of three parts: recycled naphtha, crude oil and sulphur. The revenue from these products are shown in Table 14.

Table 14: Summary of annual revenue.

Name	Unit Price	Annual Revenue (CAD)
Crude Oil	475.83 CAD/m ³	\$ 3,068,692,528.33
Naphtha	427.66 CAD/m ³	923804667.03
Sulphur	167.5 CAD/m ³	\$ 25,539,584.59
Total		\$ 4,018,036,780

The average price of WTI Cushing over 2019 was used as the estimated price of the product crude oil. WTI Cushing was selected due to its closeness in composition, as shown in Tables 15 16. It is noted that the proposed design's product has a higher weight percent of distillate and gas oil compared to the WTI Cushing. The °API gravity is converted from density and is calculated using Equation 1.1 from Gray (2015) [5]. The TR Solutions product has a higher °API density than that of the WTI product, indicating that a higher quality. This comparison shows that WTI Cushing is lower quality product, and therefore acts as a conservative estimate for potential product pricing.

Table 15: Product quality comparison.

Product Group	Cut Temp (°C)	Proposed Design (wt.%)	WTI Cushing (wt.%)
Naphtha	50-170	21.9	16.33
Distillate	170-350	40.2	42.61
Gas Oil	350-550	35.7	25.08
Residual	550-EBP	1.1	15.99

Table 16: Key parameters comparison.

Parameter	TR Solution	WTI Cushing
°API Gravity	60.5	41.5
Sulfur Content (wt.%)	1.68	0.40

The naphtha product sold is the diluent recovered from the DRU, and it is sold at a premium price of \$427.66 CAD/m³ [27]. This is \$103.66 CAD/m³ higher than the estimated diluent price (of \$324 CAD/m³).

Sulphur is assumed to be recovered from the Claus plant, as pure, solid sulphur. The market price for sulphur of \$125 USD/t was used in determining the revenue.

7.6 Cash Flow Analysis

Detailed cash flow analysis can be found in Appendix I. The project was studied under a 5 year of construction and a 35-year cash flow analysis. The parameters used for the generation of the discounted cash flow is shown in Table 17. For every year of the plant lifetime, the annual manufacturing cost is higher than the annual revenue resulting to a negative cash flow. The taxable income from this facility, also negative, was calculated as a potential taxable income credit. The negative cash flow rates indicate that the project is not economically feasible and results in a negative payback period and negative rate of return. At the end of the year 35, the net annual profit after tax (in 2020) is -\$413 Million CAD, and the NPV for this project is -\$6.94 billion.

Table 17: Parameters used in cash flow calculations.

Factors	Values
Total Fixed Capital	\$6,913,450,594
Working Capital	\$1,037,017,589
Total Expenses	\$4,583,719,631
Total Revenue	\$4,018,036,780
Tax Rate	27.00%
Discount Rate	15.00%
Inflation Rate	3.00%

7.7 Sensitivity Analysis

A sensitivity analysis is shown in Figure 6. The effect of key values changes including discount rate, capital costs, WTI Cushing price and bitumen price changes is illustrated in this figure. Details on how these key values have been changed to test the resultant NPV is in Table 18.

Table 18: Key values changing scheme on sensitivity analysis.

	Bitumen	WTI Cushing	Capital	Discount Rate
Increase	20%	20%	50%	at 20%
Decrease	-20%	-20%	-20%	at 10%

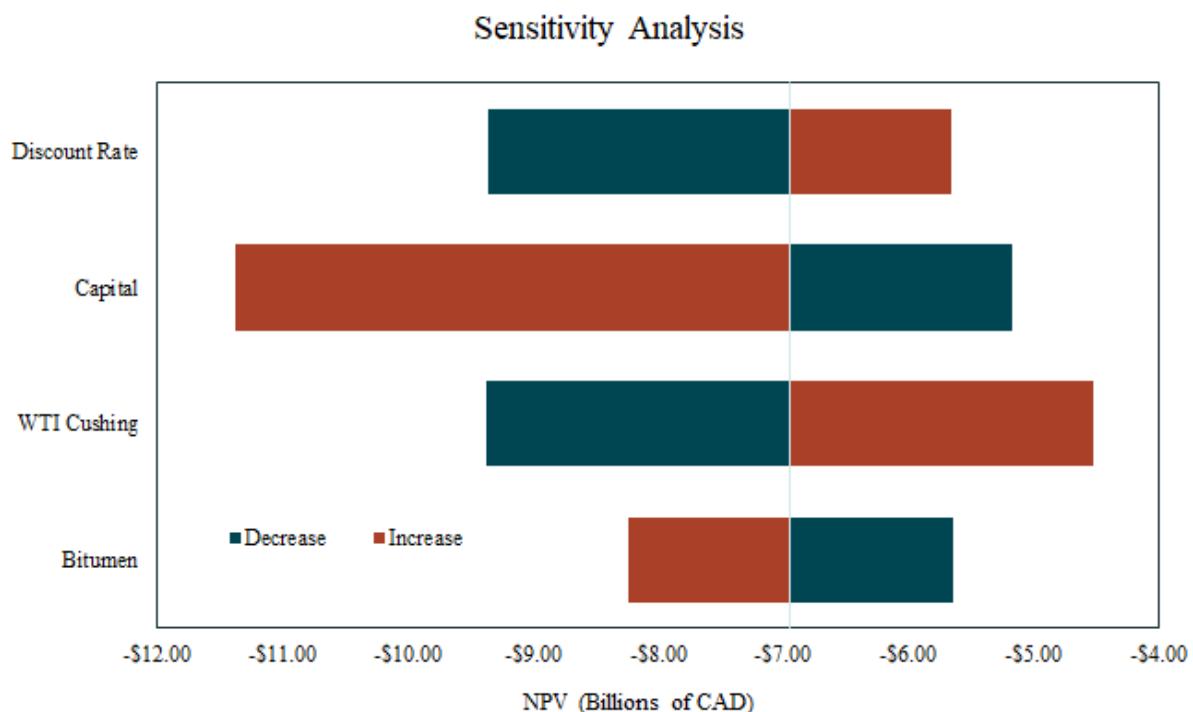


Figure 6: Sensitivity analysis.

From the sensitivity analysis, the NPV is most sensitive to the crude oil product price. If the product could be priced at 120% of its original price of \$571 CAD/m³, the NPV can rise to \$-4.52 billion CAD, and the discounted annual cash flow becomes \$400 million CAD. Furthermore, if the product price is 571 CAD/m³ or \$90.77/bbl the POT becomes 11.1 years, which is a much more reasonable number. The second largest factor that can affect the NPV is the total capital cost. A 50% increase in the capital cost can bring the NPV to as low as \$ -11.37 billion CAD. The high

sensitivity of total capital cost show that the optimization in fixed capital cost cause economic benefit to the project.

7.8 Cost of Service

The cash flow analysis concludes that the slurry hydrocracker process is not economically feasible. Table 16 shows the price sets for each feedstock/product that are required to achieve a NPV of 0; the cost of service (Church, 464 Notes) at the end of the project timeline. The feed or product price is based on average 2019 prices while the other is iteratively calculated to achieve NPV 0 at year 35. The results of this analysis are shown in Table 19.

Table 19: Cost of Service and Economic Cost/m³.

Component	Required Price (CAD/bbl)
Bitumen	\$14
Crude Oil	\$118

The above price sets are not realistic when compared to average 2019 prices. The average price of implied bitumen without condensate is approximately \$50 CAD/bbl^[34] while high quality crude sells for \$75 CAD/bbl^[34]. This is supportive evidence that the project is not economically feasible with current market conditions.

7.9 Hydrogen Option

The hydrogen production incurs high capital cost and operating cost evident from the cost estimation in sections above. Optimization and reduction of cost in hydrogen service can cause economic benefit to the overall proposed design. One way to reduce the cost is to purchase hydrogen from other sources as opposed to constructing a hydrogen plant. Appendix K shows the calculation of NPV at different purchased hydrogen options at \$3 CAD per standard cubic feet of hydrogen price. A test of purchasing 50% and 100% of the hydrogen from other sources is done, the result is in Table 20.

Table 20: Comparison of NPV by purchasing hydrogen from other sources.

% of Hydrogen Purchased	Net Annual Profit After Tax (CAD)	Return on Capital	NPV (CAD)
0% Purchase	\$-413 M	-7.79%	\$-6.94 B
50% Purchase	\$-289 M	-6.32%	\$-5.63 B
100% Purchase	\$-67 M	-2.11%	\$-3.18 B

The values above show that the cost of hydrogen production has significant effect to the net present value of the project. It is advisable to purchase hydrogen; however, it may not be available in high amounts required by an SHC unit commercially. A potential and more favourable setting for an SHC unit is an existing facility with a hydrogen plant that can increase hydrogen production to the required threshold. It can be inferred that in a fully brownfield setting, the SHC may be economically feasible if constructed with all auxiliary facilities ISBL. Further economic analyses should be conducted if the project is deemed to proceed to the next design stage.

8. ENVIRONMENTAL ANALYSIS

An environmental analysis was performed to examine the impacts of the project on the environment in terms of both direct and indirect carbon footprint. The three sources of emissions in the process are the hydrogen production process, the H₂S removal process, and the natural gas fuel. Off gas produced from the PSA unit contains H₂ gas, CO, CO₂, H₂O vapor, and some hydrocarbon gases. Combustible components in the off gas are utilized to fuel the SMR reactions, where they are converted to equivalent moles of CO₂ and H₂O vapor and emitted. The rest of energy required by the SMR as well as by the other parts of the plant is provided by natural gas fuel, which is composed mostly of methane. At the Claus plant, the sour gas is burnt in a furnace before the Claus reactors. Traces of NH₃ and hydrocarbons are converted to H₂O vapor, CO₂ and N₂ gas before being emitted. Based on the assumption that the Claus plant is equipped with a tail gas clean-up unit that will capture and recycle any unreacted H₂S, the Claus plant and the zinc oxide bed together are expected to achieve 100% H₂S removal, such that no H₂S will be released to the atmosphere. Instead, sulfur is removed from the sour gas solely in the form of elemental sulfur, S₈, which can be sold to recover value.

Local GHG emission limit has been met. According to the Oil Sands Emissions Limit Act of the Province of Alberta, the total amount of greenhouse gases (GHG) emitted from an upgrader

facility must not exceed 10 megatons per year^[17]. As shown in the sample calculations in Appendix F, this project has a total amount of 4.77 megatons of direct GHG emission annually, where 66.7% is due to PSA off gas combustion, 2.6% comes from the Claus furnace, and 30.8% is the result of natural gas burning. In addition, the project requires 6.6×10^8 kWh of electricity consumption per year, which is equivalent to 0.52 megatons of indirect CO₂ emission annually. The total carbon footprint of this project sums up to be 5.29 megatons per year. A CO₂ sequestration can potentially reduce emissions and recover some economic value. However, due to time limitations, it is not included in the scope of this design proposal.

Although natural gas is composed mainly of methane, the combustion of natural gas also produces traces of CO, NO_x, SO₂, and particulates. Based on a typical composition of natural gas, the amounts of these undesired products being released are well below any threshold for good-quality air as outlined by the Canadian Ambient Air Quality Standards, therefore are considered negligible^[23]. Overall, it is concluded that this project will not pose any significant environmental concerns.

9. SAFETY & RISK ANALYSIS

A Process Hazard Assessment (PHA) was conducted to evaluate the safety and reliability of the process. A What-if analysis was performed to examine any potential incidents that could occur at a point in the process. The Dow Fire and Explosion Index as well as the Dow Chemical Exposure Index were used to assess the explosion radii and possible chemical release. The detailed analyses are shown in Appendix G. A piping and instrument diagram (P&ID) of the three-phase separator, D-03, is illustrated in Appendix H, along with a detailed Hazard and Operability Study (HAZOP) around the unit. In each scenario, safeguards were applied to mitigate the risks to low or medium-low level. For the cases with higher than low level residual risks, recommendations were provided to further control the risks.

9.1 What-if Analyses

The PHA What if analysis was performed using the process PFD and risk matrix. The risk matrix evaluates the consequence categories of damage to asset, health and safety/ environmental risk evaluations, and the probability of an event occurring. After the causes were evaluated and the possible consequences are ranked based on the level of severity and likelihood then engineering

safeguards action are determined to bring the risk to an acceptable level according to the risk matrix. The details of the analysis, methodologies and assumptions can be found in Appendix G.

For the SHC unit the main risk studied was leakage at the different equipment of the process flow resulting in pipe failures, chemical releases, severe equipment damage and possibly fire and explosion. To mitigate those risks, it is recommended to install pressure safety valves (PSV) monitors, pressure relief valves, liquid hold-up monitors, flow control valves, regular maintenance and emergency shutdown. The slurry bubble reactor operates at very high temperature and pressure conditions which makes it prone to corrosion and damage of the column. The safeguards in place to mitigate these risks include proper insulation and pipelines padding, regular maintenance, and personal protective clothing (PPE).

9.2 Dow Fire & Explosion Index

The Dow fire and explosion classification provides a comprehensive and realistic risk analysis of certain process unit loss in the case of a rupture leading to a fire or explosion. The details of the analysis, sample calculations, methodologies and assumptions can be found in Appendix G.

Table 21: FEI, radius and area of exposure for select process equipment.

Equipment	Fire & Explosion Index	Radius of Exposure (m)	Area of Exposure (m ²)
P-01	111	27.4	2359
F-01	188	48.8	7482
F-02	132	33.5	3526
D-01	95	24.4	1870
D-02	95	24.4	1870
D-03	125	32.0	3217
R-01	139	35.7	4004
C-01B	140	36.6	4208
E-02	145	37.8	4489

For this project analysis nine process units were selected to be analysed, including feed pump, bitumen furnace, hydrogen furnace, two overhead flash drums, three-phase separator, slurry hydrocracker reactor and two heat exchangers. Table 21 shows the fire and explosion indices for each process unit mentioned above, as well as the radius and area of exposure.

Due to the high operating conditions in equipment such as reactors trains (R-01A to R-04B), furnace (F-01 and F-02), and heat exchangers (C-01B and E-2) the area of exposure was considerable large ranging from 3526 m to 7482 m compared to the flash drums. Therefore, separating the equipment considering the FEI indices and area of exposure is crucial to mitigate the risk of fire or fuel-air explosion escalation in the case of an incident, as well as, adequate PPE and engineering and administrative controls.

9.3 Dow Chemical Exposure Index

The chemical exposure index is the risk analysis related to the health hazards associated with chemical releases, it estimates the airborne quantity released and rates the potential health hazard to neighbouring plants and communities in the event of chemical release incident. The details related to the CEI analysis including, sample calculations, methodologies and assumptions can be found in Appendix G.

Table 22: CEI, hazard distance and release duration chemical release.

Case	Type of Release	Chemical Released	CEI	Hazard Distance (m)	Maximum Release Duration (min)
1	Piping	Vapor H ₂ S	939	9395	5
	Piping	Liquid Ammonia	597	5970	6
2	Piping	Vapor Ammonia	46	461	5
	Piping	Vapor H ₂ S	86	863	11

Table 22 shows the chemical exposure index calculated for two different cases in the process, as well as the hazard distance and chemical release duration. The two cases consisted of pipe leaks involving different toxic chemicals. Case 1 has a significant high CEI due to the large amount of vapor hydrogen sulfide and liquid ammonia present in the pipe. For case 2 vapor hydrogen sulfide and ammonia are present in the pipe and the CEI is lower, regardless it still requires control mitigations. Due to the high indices and release duration proper selection of materials for piping is needed to reduce the risk of a leak as well as safety valves, level flow control valves, and emergency shutdown in case a leak occurs.

9.4 HAZOP

A HAZOP study was performed around the three-phase separator unit, D-03. In Appendix H, a P&ID of the unit is illustrated as well as the detailed hazards and safe controls.

Potential hazards of the unit are mainly due to deviations in temperature, pressure, and liquid levels in the drum. In case of an abnormal feed temperature, temperature indicators combined with a variable frequency drive (VFD) control the cooling rate provided by the air cooler to maintain a steady feed temperature. Liquid levels accumulated inside the drum are monitored by level indicators, and signals are sent to liquid control valves on the liquid draw streams, which respond with direct actions to the liquid levels. Under the circumstance of elevated pressures inside the vessel, a pressure safety valve (PSV) responds with releasing materials to the flare in order to reduce the pressure back to the normal range. A pressure indicator coupled with a pressure control valve being installed on the gas-phase outlet can also adjust the valve position to maintain a consistent pressure of gas being sent downstream.

The analysis result shows that the existing safety controls have reduced the risk levels of all of these scenarios to either low or medium-low levels. Nevertheless, since the high pressure condition can lead to severe consequences in case of fire or explosion, it is recommended that all workers on site must wear PPE all the time, and management must provide the workers with adequate training, accessible procedures, and an evacuation plan.

9.5 Plot Plan

The plot plan of the facility is illustrated in Appendix D. The positioning of the major equipment was based on a counter-clockwise direction of process flow. Equipment placement was made by spacing units according to the Dow Fire and Explosion radius of exposure. Equipment with the most significant radii are marked with red circles.

The control room is located at the northwest corner of the area, where it is kept away and up-wind of the most hazardous equipment.

10. NET SOCIAL BENEFIT

The net social benefit of the project will be analysed to study the overall impact of the facility on the local communities. The location of the facility, Edmonton, is populated with one million residents^[20]. Several impacts can be made by the construction of this project on greenfield. Noises caused by the processing can have a potential negative impact on the community. Another

negative social impact of the facility is the environmental footprint due to emissions. The aesthetics and traffic of the area can be affected during the construction of the plant. From the perspective of economics, the process will provide jobs and promote the economy to the Alberta region. It will explore and popularize an advanced technology of bitumen upgrading in comparison to the traditional coking method. Higher conversion of the products will be achieved, and accessible synthetic crude oil will be available in the market.

11. CONCLUSIONS & RECOMMENDATIONS

11.1 Conclusions

The proposed design allows processing of 100,000 equivalent barrels of bitumen per day and produces 123,500 bbl/day of premium crude product with Naphtha, Light Gas Oil and Heavy Gas Oil distributions of 21.9 wt.%, 40.2 wt.% and 35.7 wt.% respectively. The proposed location of the facility is located in Edmonton, Alberta. The intended purpose of this project is to analyse the economic feasibility and attractiveness of the proposed design that is legislation compliant. The proposed design is a greenfield standalone site that utilizes the slurry phase hydrocracking technology with a steam reformer unit, amine treater unit and a Claus plant to produce high quality crude.

The estimated erected capital cost of the entire facility is \$6.9 billion CAD. The annual manufacturing expense of the facility including indirect expenses, raw material feeds, and utility costs is \$3.3 billion CAD. The product streams including product oil, naphtha and sulphur generate a revenue of \$4.0 billion CAD a year. The 35-year discounted cash flow analysis showed the payback period is 997 years with a rate of return of 0.10% within the project lifetime which show that this project is not economically feasible. Multiple evidences and analyses show that the cost of hydrogen including the capital cost and operating cost is the main factor of the project's infeasibility. There is also not a sufficient profit margin from high quality crude and bitumen to support the annual manufacturing expense of the facility.

From the perspective of safety and risks, this project does not pose significant concerns. The local environmental regulation has been complied with. Safety controls are applied such that all residual risks of potential safety concerns are of low or medium-low levels as well.

11.2 Recommendations

Since the proposed project is not economically feasible, TR Solutions does not recommend the proceed to the next step of the design. Following recommendations can be explored further to analyse the viability of the facility.

11.2.1 Crude Fractionation and Brownfield Option

The economic analysis conducted for this project showed that the margin between bitumen and high-quality crude is not sufficient for the feasibility of a slurry hydrocracker unit. Therefore, it is noteworthy to further explore the feasibility of this unit with a crude fractionation along with hydrotreating refining units to produce end-user petroleum products such as gasoline, diesel and jet fuel to leverage higher profit margins. The slurry hydrocracker unit can also be a potential front-end unit prior to the crude distillation in a refinery allowing the facility to process heavier crude oils and diluted bitumen. A brownfield project would inquire less capital cost since the auxiliary facilities are available.

11.2.2 Heat Integration

A formal heat integration pinch analysis can be conducted to find the most efficient heat exchanger network configuration.

11.2.3 CO₂ Sequestration

The proposed design produces 3.3 MT of greenhouse gas emissions annually. Sequestering these greenhouse emissions is a viable option for environmental stewardship. the produced emissions can inquire for financial credit. The vent gas from the SMR can be routed to the Amine unit as an amine-based sequestering.

11.2.4 Purchased Hydrogen Option

TR Solutions recommend exploring the project in scenarios with the potential to purchase process hydrogen. The economic analysis above concludes that the project is feasible at full purchase capability for hydrogen. The resultant NPV is \$-1.14 billion CAD with a payback period of 9.7 years.

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SLURRY HYDROCRACKER PROJECT

Appendix A - Mass & Energy Balance

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A.1 SUMMARY

Appendix A contains detailed lists of mass and energy balances for the overall process, as well as for the bubble column reactor. Specific balances are provided for the reactor to ensure correct operation given that the reactor is not modelled in the simulation.

Included in the supplementary files are two Excel spreadsheets with detailed calculations for mass balances, energy balances, and product distribution, as well as an Excel copy of the stream table.

A.2 MASS BALANCE

A.2.1 Overall Process Mass Balance

Table A1 shows the mass balance for the entire process. The total difference was found to be 0.24 kg/h, or 0.00003%.

Table A1. Overall process mass balance.

In			Out		
Stream Description	Stream Number	m (kg/h)	Stream Description	Stream Number	m (kg/h)
Bitumen Feed	S1	665,516	Final Product	S60	602,327
Steam	S26	139,000	Residue Product	S45	14,424
Methane	S27	39,845	SMR Liquid	SMR-S09	1,085
Stripper Steam	S41	4,500	Claus Plant	CP-S01	23,797
Fresh Catalyst	S7	6,655	PSA Purge	S32	207,918
Combustion Air	-	1065	Waste-Water	S46	7,029
Total:		856,581	Total:		856,581

A.2.2 Reactor Mass Balance

Table A2 shows the combined mass balance around the slurry bubble column reactors (with specific product composition).

Table A2. Reactor Mass Balance.

In		Out	
Stream Description	m (kg/h)	Stream Description	m (kg/h)
Bitumen Feed	665,516	Product Gas	91,369
(Fresh) Catalyst	6,655	Naphtha	128,937
Fresh H ₂	40,176	LGO	242,047
Recycle Gas	140,052	VGO	215,123
		Residue	14,716
		(Spent) Catalyst	6,655
		Unreacted H ₂	13,500
		Recycle Gas	140,052
Total:	852,400	Total:	852,399

A.3 ENERGY BALANCE

A.3.1 Overall Energy Balance

A summary table for the overall process energy balance is presented in the following table. It was assumed that all reactors operate isothermally. Stream enthalpies were taken directly from VMG Symmetry and can appear positive or negative based on the arbitrary enthalpy baseline set by VMG.

Table A4. Overall process energy balance.

Stream Description	Stream Number	Electricity Input (kW)	Heat In (kW)	Heat Out (kW)	Energy In (kW)	Energy Out (kW)
Bitumen Feed	S3	-	-	-	-11,814	
Steam	S33	-	-	-	39,592	
Methane	S34	-	-	-	10,800	
Stripper Steam	S44	-	-	-	862	
Final Product	S60	-	-	-		3,165
Residue Product	S45	-	-	-		430
SMR Liquid	SMR-S09	-	-	-	506	
Acid Gas	S30	-	-	-	712	
PSA Purge	S35	-	-	-	30,559	
Waste-Water	S52	-	-	-	3,282	
Equipment Description	Equipment Number	Electricity Input (kW)	Heat In (kW)	Heat Out (kW)	Energy In (kW)	Energy Out (kW)
Slurry Bubble Column Reactor(s)	R-01 – R-04	-	93,536	-	93,536	
Conversion Reactor	R-05	-	-	8,151		8,151
Pump Feed Train	P-01	4,965	-	-	4,965	
Final Product Pump	P-02	1	-	-	1	
Compressor Train Compressors	CT-CP01 / CT-CP02	54,734	-	-	54,734	
Compressor Train Cooler	CT-C01	-	-	26,753		26,753
Product Gas Expander	EX-01	-	-	-		4,642
Product Overheads Cooler	C-01	-	-	211,564	211,564	
Product Bottoms Cooler	C-02	-	-	1,030		1,030
Stripper Cooler	C-03	-	-	4,559		4,559
3-Phase Separator Cooler	C-04	-	-	41,843		41,843
Bitumen Feed Furnace	F-01	-	57,014	-	57,014	
Internal Recycle Gas Furnace	F-02	-	56,431	-	56,431	
Bitumen Feed Pre-heater	E-01	-	64,949	-		
Hydrogen Pre-Heater	E-02	-	15,038	-		
Internal Recycle Gas Compressor	CP-01	480	-	-	480	
Stripping Unit Overheads Compressor	CP-02	9,581	-	-	9,581	
SMR/PSA	-	-	-	-	12,924	
Amine Unit	AU	-	-	-		3,115
					Total:	329,105
					Difference:	-2,204
						1%

A.3.2 Reactor Energy Balance

Table A5 shows the overall energy balance for the (combined) slurry bubble column reactors. The table shows that the overall difference is only ~1%, and is therefore acceptable.

Table A5. Slurry bubble column reactor energy balance.

Stream Description	Stream Number	Electricity Input (kW)	Heat In (kW)	Heat Out (kW)	Energy In (kW)	Energy Out (kW)
Reactor Product	"	-	-	-	324,341	
Vapour Overheads	S11	-	-	-		419,438
Liquid Bottoms	S10	-	-	-		640
Equipment Description	Equipment Number	Electricity Input (kW)	Heat In (kW)	Heat Out (kW)	Energy In (kW)	Energy Out (kW)
Slurry Bubble Column Reactor(s)	R-01 – R-04	-	93,536	-	93,536	
					Total:	417,877
					Difference:	2200
					Equipment Total:	420,077
					Difference:	1%
					Equipment Total:	93,536
					Difference:	0
					Equipment Total:	93,536
					Difference:	-

A.4 PRODUCT DISTRIBUTION

Data here is provided to show a comparative analysis between different liquid products produced, and to compare final product composition with (Athabasca bitumen) feedstock.

A.4.1 Liquid Product Distribution

Table A3 shows a comparative component yield analysis for three liquid hydrocarbon products; the original reactor liquid (before being treated in the stripping process), the final (saleable) product, as well as the catalyst-containing residue product. Data includes stream composition both with and without spent catalyst.

Table A3. Component distribution for various liquid products.

Boiling Cut / Component	Cut Temperature	Weight % (w/o Catalyst)	Weight % (w/ Catalyst)
Original Reactor Liquid			
Lights	IBP - 50	0.0	0.0
Naphtha	50-170	21.5	21.2
Light Gas Oil	170-350	40.3	39.8
Heavy Gas Oil	350-550	35.8	35.4
Residue	550 - EBP	2.4	2.4
Spent Catalyst	N/A	-	1.1
Final Product			
Lights	IBP - 50	2.9	2.9
Naphtha	50-170	20.1	20.1
Light Gas Oil	170-350	40.2	40.2
Heavy Gas Oil	350-550	35.6	35.6
Residue	550 - EBP	1.2	1.2
Residue Product			
Lights	IBP - 50	0	0.0
Naphtha	50-170	0	0.0
Light Gas Oil	170-350	0	0.0
Heavy Gas Oil	350-550	5.5	3.0
Residue	550 - EBP	94.5	50.9
Spent Catalyst	N/A	-	46.1

A.4.2 Final Product-Feedstock Comparisons

Figure A1 displays a distillation curve of the final product compared with the distillation curve of pure bitumen from Gray (2015).

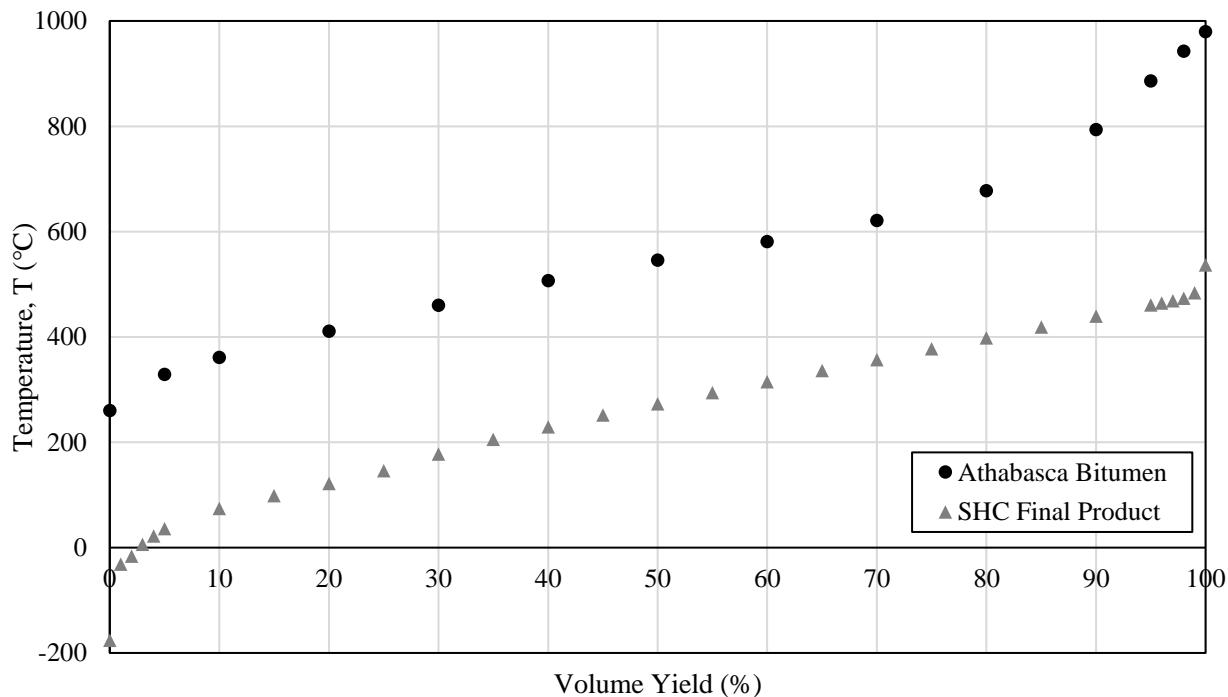


Figure A1. Distillation curves for Athabasca bitumen (feed) and SHC final product.

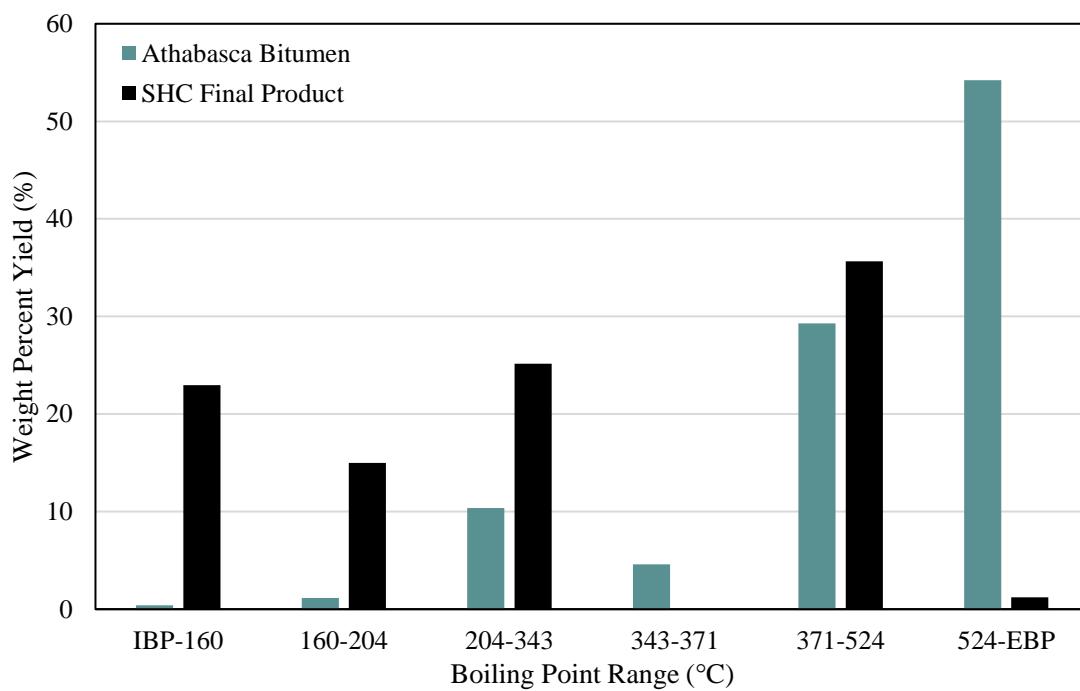


Figure A2. Composition of boiling fractions for Athabasca bitumen and SHC final product.

A.5 REFERENCES

- Ghasemi, M. & Whitson, C. (2013). Modeling Steam-Assisted Gravity Drainage With a Black-Oil Proxy. *SPE Reservoir Evaluation & Engineering*. 16, 155-171. 10.2118/147072-PA.
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SLURRY HYDROCRACKER PROJECT

Appendix B - VMG Simulation

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B.1 SUMMARY

This Appendix contains all files and information related to the development/design and utilization of the VMG Symmetry simulation of the proposed design. Included files/documents relating to the simulation include current outputs, current and previous simulation files, and a supplementary simulation file for amine unit model estimations.

A total of 50 iterations were completed in simulation development. The thermodynamic model used was the refinery characterization of Advanced Peng Robinson. Simulation development was done through a review of current refinery processing and assistance from our academic advisor.

B.2 SIMULATION OUTPUTS

A copy of the simulation PFD is provided below. A copy of the VMG Symmetry-generated simulation report is included with the simulation supporting files.

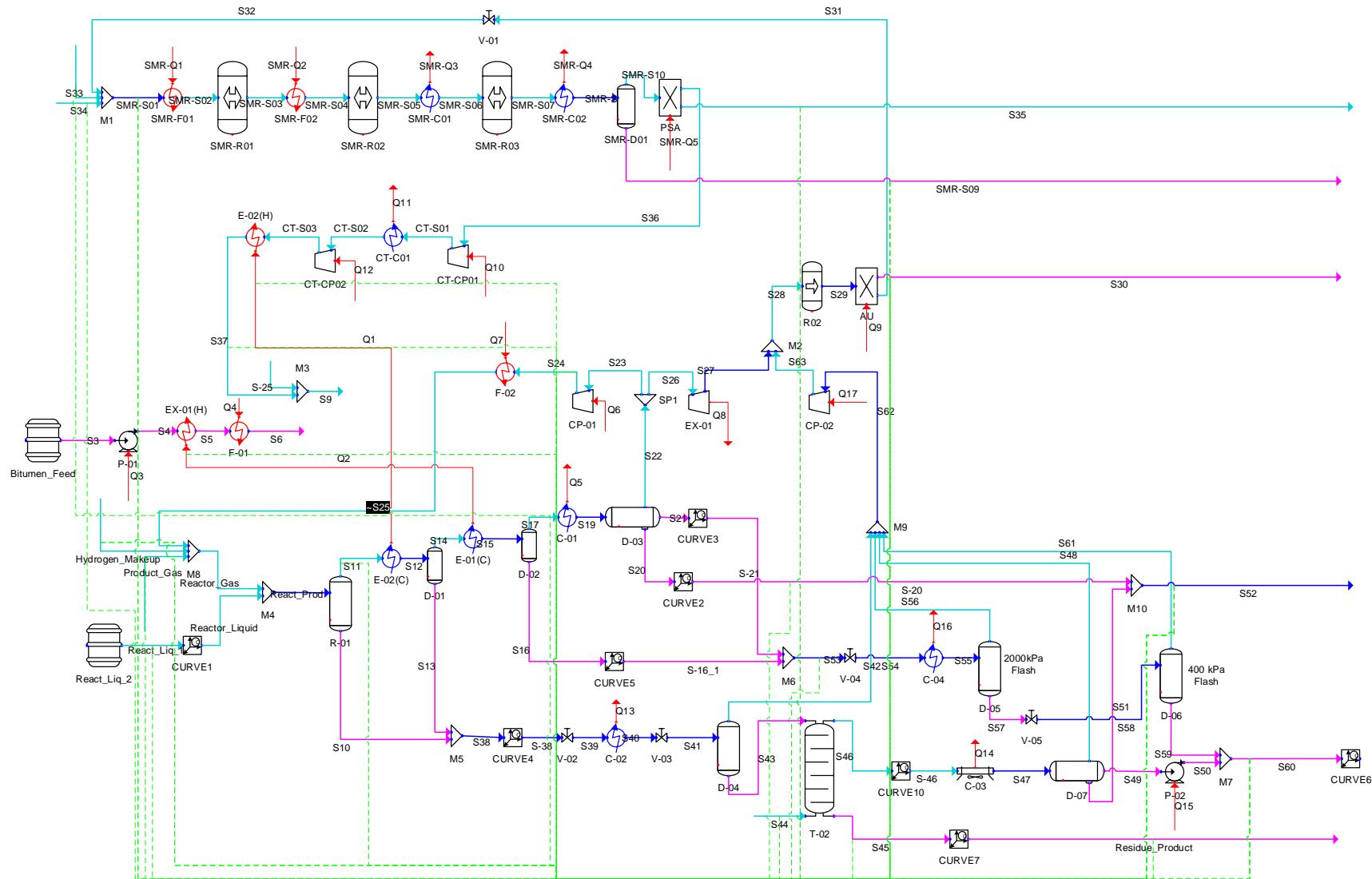


Figure B1. Simulation Process Flow Diagram

B.3 SIMULATION COMPONENTS

Certain simulation components are either not simulated or have not undergone detailed design (due to them being outside the scope of our project).

B.3.1 Bubble Column Reactor

Since the proposed hydrocracker design was based on an existing process, final product composition (using the catalyst of choice) was known, and therefore did not require reactor modelling to predict. As a result, no slurry bubble column reactor was modelled in Symmetry. Instead, the known CANMET product data was characterized in VMG with pseudo components (based on a carbon number distribution). In lieu of a modelled reactor, known liquid and gaseous product streams were mixed in a combined reactor product stream and placed into a two-phase separator to re-create the separation that would occur in the real slurry bubble column reactor.

It was found that most of the liquid product vaporized in the separator, and in conferring with our academic advisor, it was determined that this would be unlikely in the real process. The consequence of this is that our overheads are likely oversized (but this of little concern).

Additionally, VMG Symmetry does not have component support for iron (II) sulfate (the chosen process catalyst), so catalyst flows were not simulated at any time. For operation, it was assumed that catalyst flows would be added to the inlet bitumen stream and would flow out of the column with the liquid product. From there it would be completely removed from the system with the final residue product.

B.3.2 Amine Unit

The amine unit, represented as the component splitter AU, was designed to function like a real amine unit running with MDEA. In this, it separates 100% of the inlet H₂S and NH₃, and pulls out a small fraction of various other components while respecting appropriate inlet and outlet temperatures. The specific separation percentages were determined by running the inlet through an external amine unit simulation (provided in ChE 464) and converting the results into separation percentages. A copy of the amine unit simulation is provided in the supplemental materials.

B.3.3 Claus Plant

Simulations were not developed for Claus plant operation due to the highly available, well-documented nature of operating data for industrial Claus plants. Values for operation were estimated from literature data.

B.3.4 Steam Methane Reformer & Pressure Swing Adsorber

Since the simulation results regarding H₂ production were significant, it was required to simulate an SMR and a PSA (despite being outside the scope of detailed engineering). Values for SMR and PSA operation were taken from literature in order to ensure inlet and outlet conditions were correct, but specific optimization was never performed for these components.

The specific SMR reactors (SMR-R01, SMR-R02, and SMR-R03) are modelled as equilibrium reactors with the respective hydrocarbon-steam-reforming and water gas shift reactions specified.

B.3.5 Internal Gas Recycle Loop

The loop is required for maintaining high partial pressure of H₂. Only basic optimization was performed during development. The target was to have exit vapour composition be 65 mol% (or greater) H₂. It was found that increasing the size of the recycle loop positively increased H₂ partial pressure but came at the cost of ridiculously high flow rates. The higher flowrates ultimately caused H₂ partial pressure to decrease in the overheads, so a balance was struck by keeping the internal recycle fixed at 70 mass% of the total vapour overheads (from the three phase separator).

B.3.6 Conversion Reactor (Hydrotreater)

Due to the presence of sulfur and nitrogen-containing compounds from the stripping system overheads, a hydrotreater was required to prevent these compounds reaching the SMR (as they would poison the catalysts). Reactor R-05 is modelled as a conversion reactor which converts large sulfur and nitrogen containing hydrocarbons (C₆ and larger) into hydrogen sulfide, ammonia, carbon dioxide and water. The unit is located just before the amine unit in order to separate out the ammonia and hydrogen sulfide.

B3.7 Product Characterization

Both bitumen feed and reactor liquid product were characterized from boiling curve data and pseudo-component selection (based on carbon number) from VMG Symmetry. Distillation curves of various hydrocarbon products are provided below.

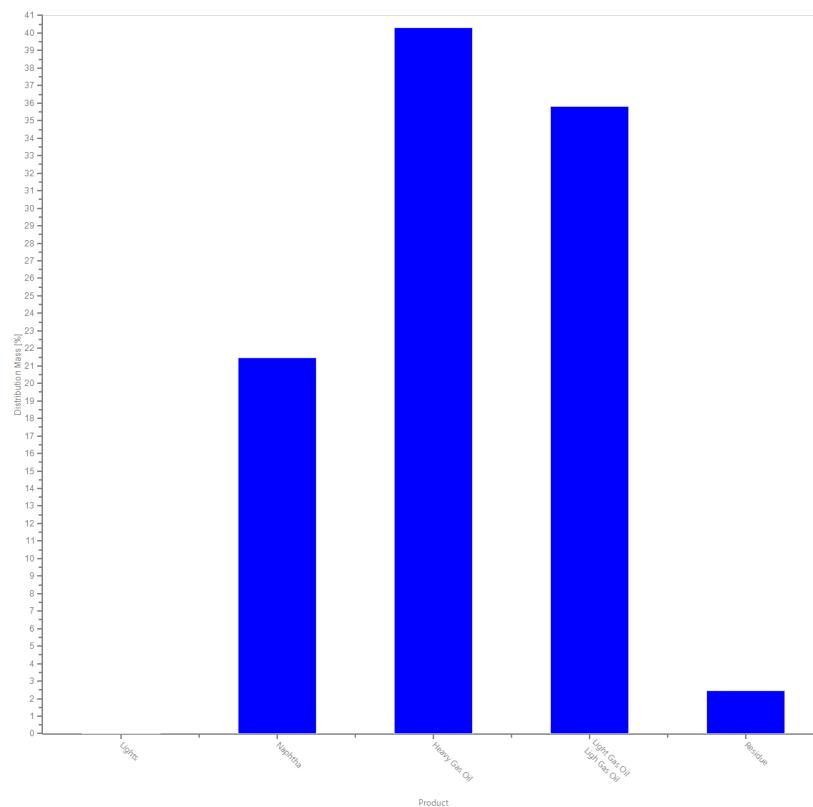


Figure B2. Product distribution of characterized reactor liquid (initial liquid product).

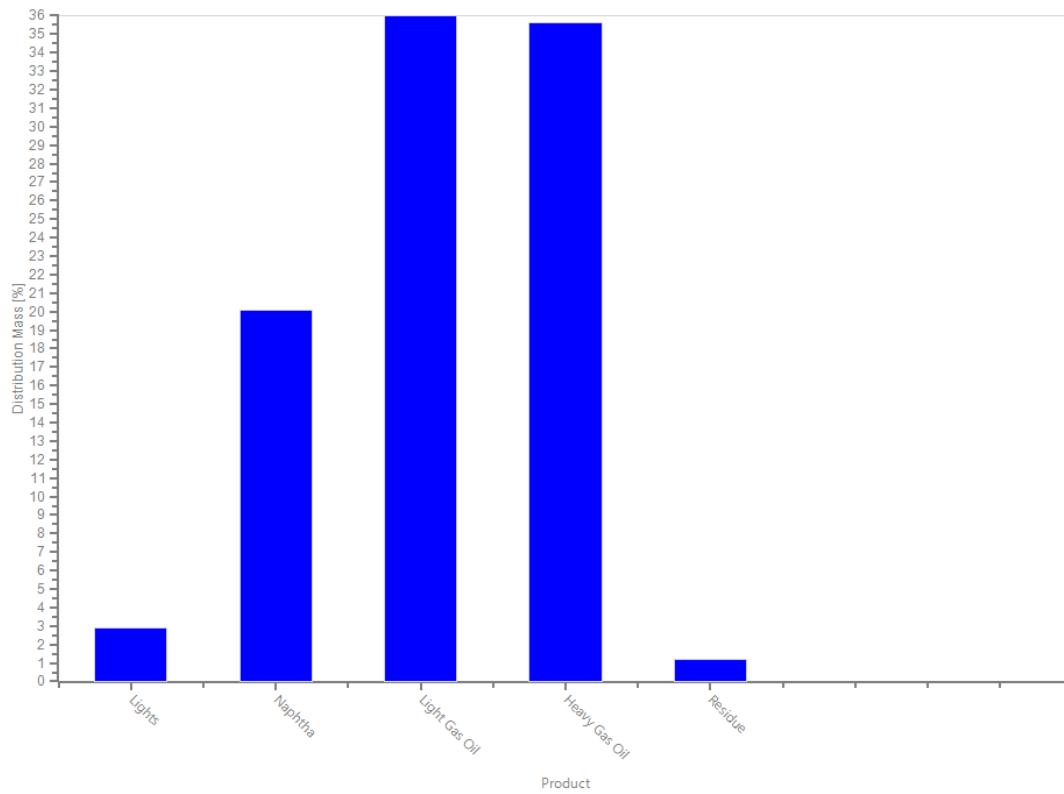


Figure B3. Product distribution of final liquid product.

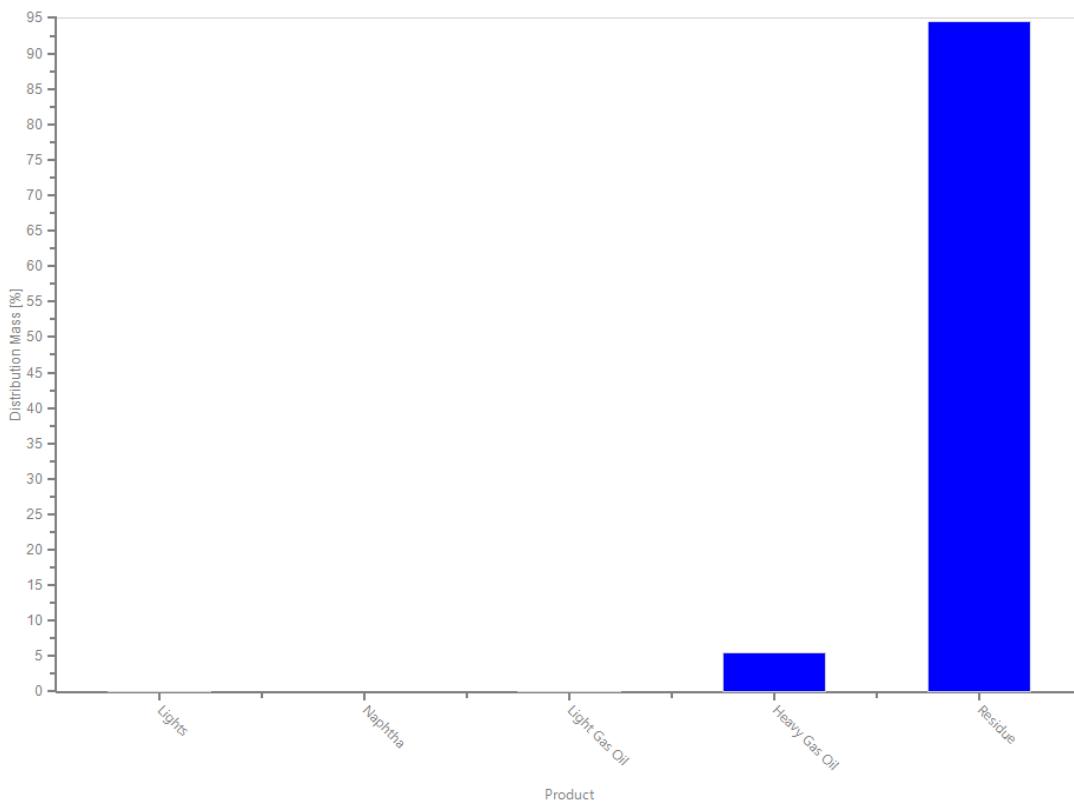


Figure B4. Product distribution of final residue product.

B.4 MODEL VALIDATION

Simulation outputs were validated by comparing a simulation product stream outlet with known material properties from literature. Results comparing simulation outputs with available data are provided in Table B1.

Table B1. Density of Athabasca bitumen vacuum reside (>524°C).

Component	Component Density (SG)
Simulation Results	1.16
Ghasemi et al., 2011	1.06
Difference:	10%

B.5 REFERENCES

- Ghasemi, M. & Whitson, C. (2013). Modeling Steam-Assisted Gravity Drainage With a Black-Oil Proxy. *SPE Reservoir Evaluation & Engineering*. 16, 155-171. 10.2118/147072-PA.

SLURRY HYDROCRACKER PROJECT

Appendix C - Reactor Mass Balance

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C1 SUMMARY

This Appendix shows the elemental mass balance around the hydrocracking reactor. The feed components are obtained from Athabasca bitumen composition, and the product yields as well as their sulphur and nitrogen contents are based on CANMET product data [1,2]. Oxygen contents in the products are arbitrary numbers based on the fact that oxygen typically takes up 0.08 - 1.8 weight % in synthetic crude oils, where it tends to be in heavier hydrocarbons [3].

C2 REACTOR COMPONENT BALANCE

Table C1. Feed and product elemental compositions.

Elements	Feed (wt.%)	CANMET Product Composition (wt.%)				
	Bitumen	Naphtha	LGO	VGO	Residue	
C	82.70%	84.10%	83.86%	83.71%	82.45%	
H	10.50%	15.12%	13.36%	12.31%	11.27%	
S	5.10%	0.60%	1.80%	2.30%	3.10%	
N	0.60%	0.10%	0.40%	0.60%	1.60%	
O	1.10%	0.08%	0.58%	1.08%	1.58%	

Table C2. CANMET product percent yields.

CANMET Yield (wt%)	
Naphtha	19.80%
LGO	33.50%
VGO	28.50%
Residue	4.50%
Gas product	13.70%

Table C3: Elemental mass flow rates of the feed and products.

Elements	Feed (kg/h)		Products (kg/h)					
	Bitumen	H2 consumed	Naphtha	LGO	VGO	residue	Gas product	
Total	665516.00	27060.47	137130.14	232013.12	197384.30	31165.94	94882.98	
C	550381.73	0	115326.10	194574.33	165236.41	25695.29	49549.59	
H	69879.18	27060.47	20734.42	30988.82	24291.99	3513.43	17410.99	
S	33941.32	0	822.78	4176.24	4539.84	966.14	23436.32	
N	3993.10	0	137.13	928.05	1184.31	498.66	1244.95	
O	7320.68	0	109.70	1345.68	2131.75	492.42	3241.12	

Table C4. Overall mass balance around the hydrocracking reactor.

Overall Mass Balance (kg/h)	
Bitumen feed	665516
H2 consumed	27060.47
Total in	720576.47
Naphtha	142674.14
LGO	241393.12
VGO	205364.30
Vacuum residue	32425.94
Gas product	70718.98
Total out	720576.47

C3 REFERENCES

- [1] Gray, M. R. (2015). *Upgrading oilsands bitumen and heavy oil*. Retrieved from <https://ebookcentral.proquest.com>.
- [2] Speight, J.G. (2017). *Handbook of petroleum refining*, CRC Press, Taylor & Francis Group.
- [3] *Robust Summary of Information on Crude Oil*. (2011).

SLURRY HYDROCRACKER PROJECT

Appendix D - Drawings

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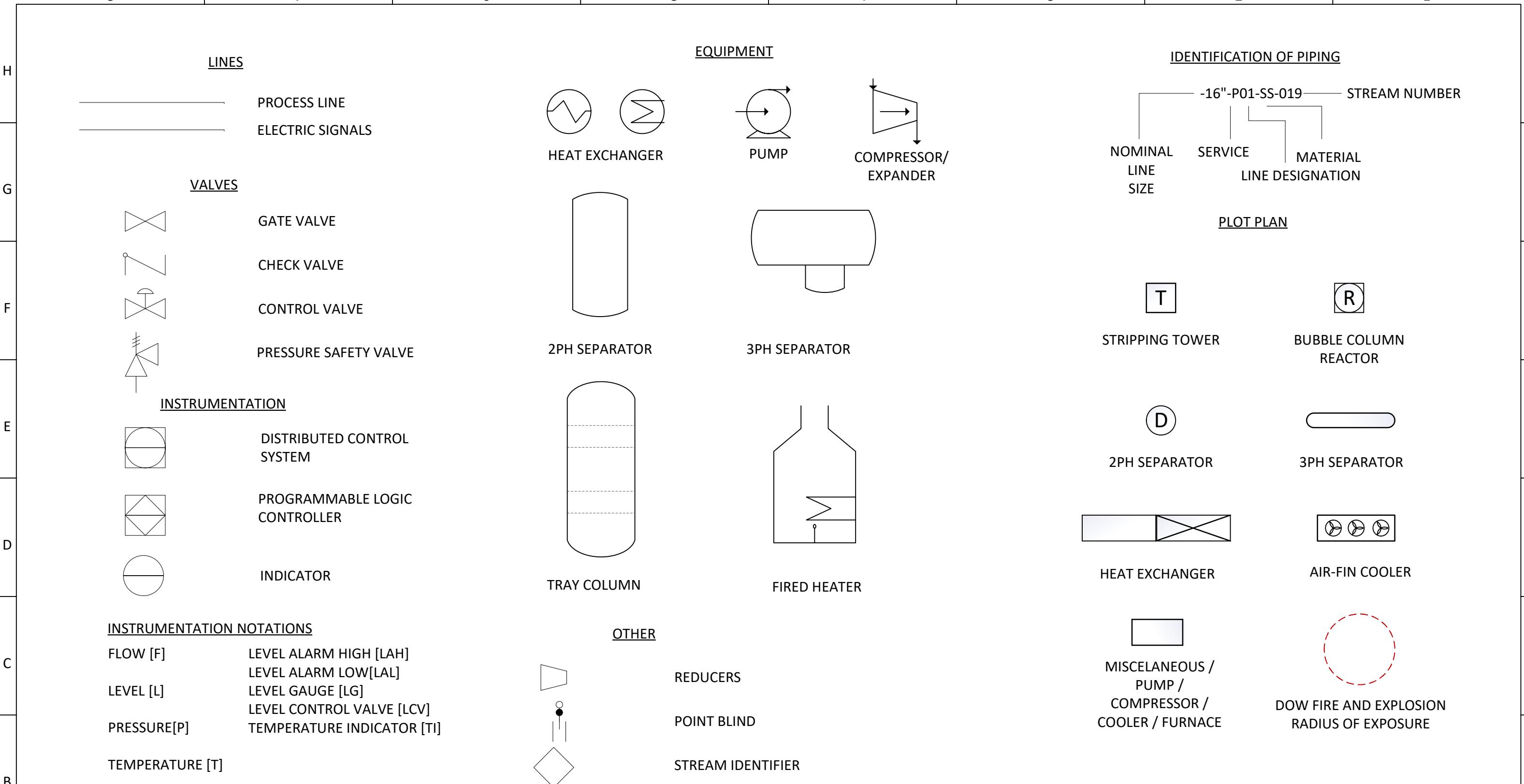
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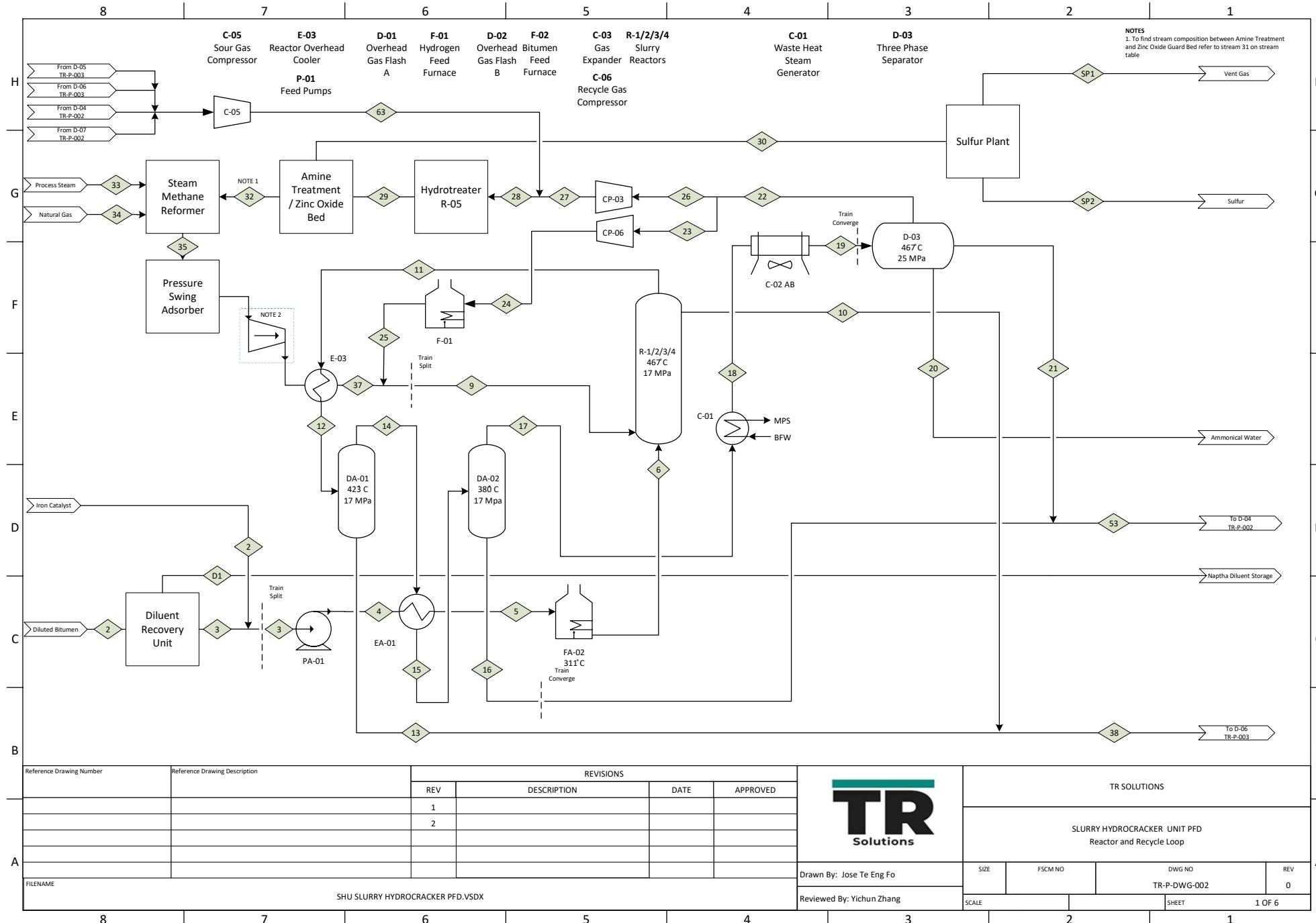
D.1 SUMMARY

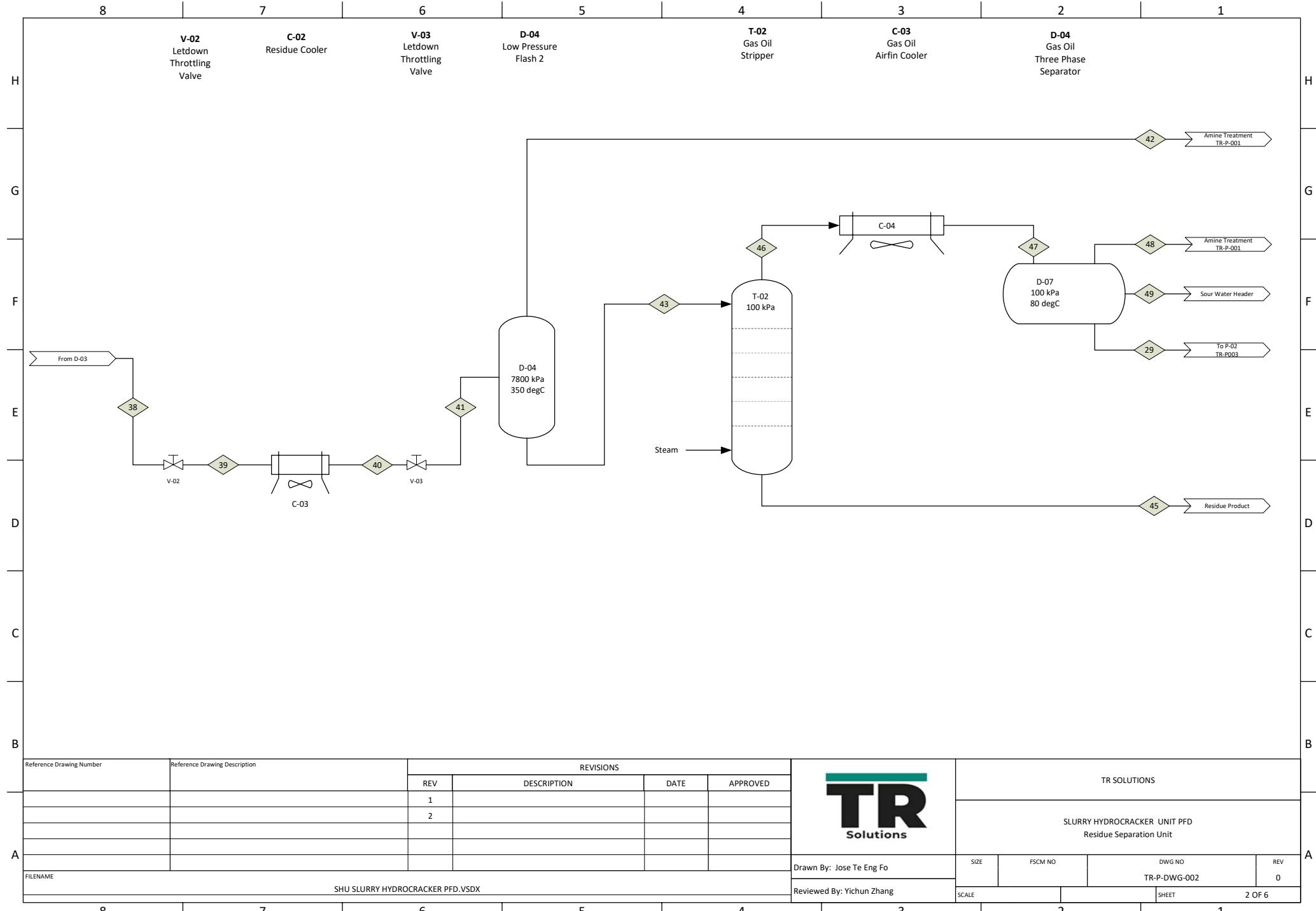
Appendix D contains the files and information with the included drawing package of the proposed facility design. The drawing standard has a complete list of symbology that are utilized across all drawings. The PFD attached outlines all process steam with the corresponding detailed stream table. A process and instrumentation diagram (P&ID) with line list is also included.

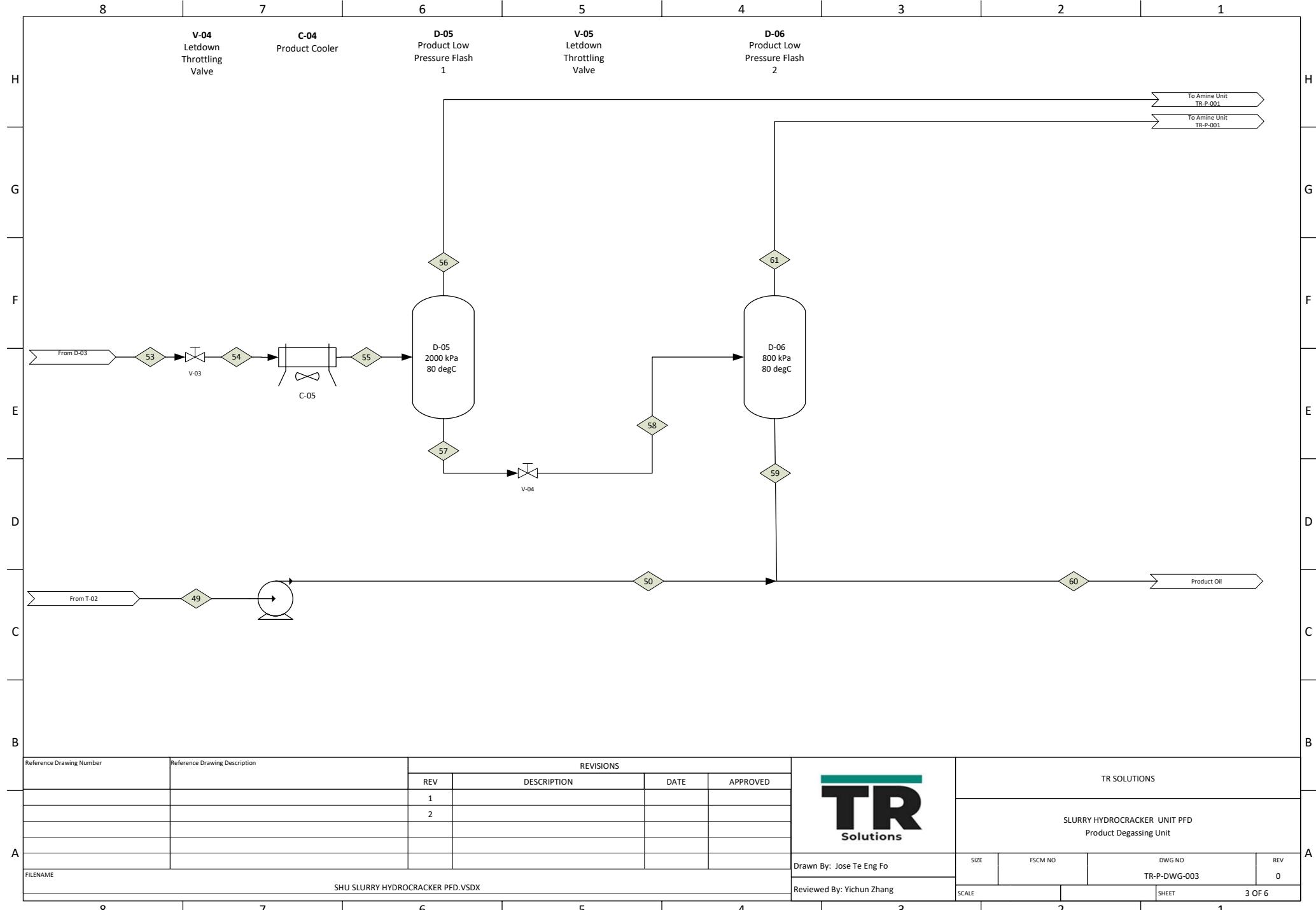
D.2 DRAWING STANDARDS / PROCESS FLOW DIAGRAM

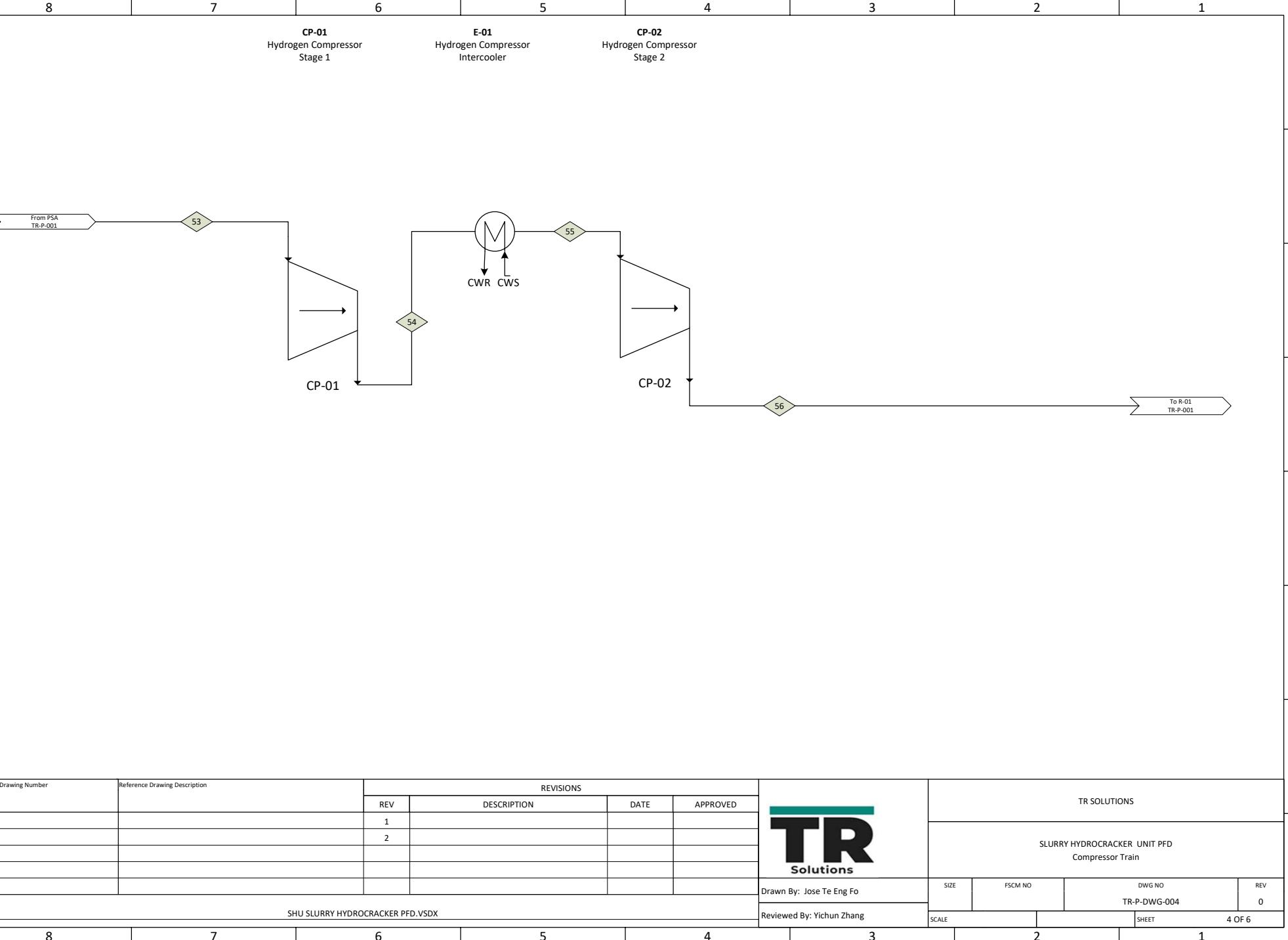
8 | 7 | 6 | 5 | 4 | 3 | 2 | 1











D.2.1 Process Stream Table

Table D1. Detailed stream table with component mass fractions.

Components	S9	S10	S11	S12	S13	S14	S15	S16
T (°C)	354.8	466.9	466.9	450.0	450.0	450.0	380.0	380.0
P (kPa)	17000.00	17000.00	17000.00	16900.00	16900.00	16900.00	16800.00	16800.00
Vapour Fraction (-)	1.00	1.00E+00	1.00	0.9990	0.00	1.00	0.9660	0.00
Mole Flow (k·mol/h)	38643.07	50.37	31985.51	31985.51	30.76	31954.75	31954.75	1085.75
Mass Flow (kg/h)	180228.8	9175.69	843223.9	843223.9	10164.10	833059.8	833060	187409.5
Mass Fraction (-)								
HYDROGEN	0.3809	2.28E-04	0.0498	0.0498	0.0011	0.0504	0.0504	0.0028
METHANE	0.2700	0.0006	0.0895	0.0895	0.0025	0.0905	0.0905	0.0066
ETHANE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
PROPANE	0.0795	0.0003	0.0357	0.0357	0.0016	0.0361	0.0361	0.0043
n-BUTANE	0.0569	0.0004	0.0368	0.0368	0.0020	0.0372	0.0372	0.0055
CARBON MONOXIDE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CARBON DIOXIDE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NITROGEN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
WATER	0.0078	4.98E-05	0.0057	0.0057	2.23E-04	0.0058	0.0058	5.37E-04
AMMONIA	0.0071	2.45E-05	0.0032	0.0032	1.12E-04	0.0033	0.0033	2.95E-04
HYDROGEN SULFIDE	0.1524	0.0006	0.0600	0.0600	0.0025	0.0607	0.0607	0.0063
IRON (II) SULFATE	0.00	0.73	0.00	0.00	0.00	0.00	0.00	0.00
SULFUR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-HEXANE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Paraffin_C6	0.0229	0.0006	0.0398	0.0398	0.0030	0.0402	0.0402	0.0085
HC_Paraffin_C7	0.0113	0.0007	0.0369	0.0369	0.0032	0.0373	0.0373	0.0092
HC_Paraffin_C8	0.0054	0.0008	0.0353	0.0353	0.0036	0.0357	0.0357	0.0103
HC_Paraffin_C9	0.0026	0.0009	0.0344	0.0344	0.0040	0.0348	0.0348	0.0117
HC_Paraffin_C10-C14	9.30E-04	0.0034	0.0933	0.0933	0.0161	0.0943	0.0943	0.0485
HC_Paraffin_C15-C19	3.93E-05	0.0106	0.1475	0.1475	0.0508	0.1487	0.1487	0.1496
HC_Paraffin_C20-C35	5.52E-08	0.0414	0.1751	0.1751	0.2024	0.1748	0.1748	0.4243
HC_Paraffin_C36+	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_C6	9.14E-04	5.86E-05	0.0028	0.0028	2.71E-04	0.0028	0.0028	7.41E-04
HC_Aromatic_C7	4.86E-04	7.47E-05	0.0030	0.0030	3.46E-04	0.0030	0.0030	9.55E-04
HC_Aromatic_C8	2.61E-04	9.61E-05	0.0032	0.0032	4.46E-04	0.0033	0.0033	0.0012
HC_Aromatic_C9	1.39E-04	1.22E-04	0.0035	0.0035	5.67E-04	0.0035	0.0035	0.0016
HC_Aromatic_C10-C14	6.93E-05	0.0007	0.0124	0.0124	0.0031	0.0126	0.0126	0.0085
HC_Aromatic_C15-C19	5.21E-06	0.0031	0.0297	0.0297	0.0147	0.0299	0.0299	0.0382
HC_Aromatic_C20-C35	2.31E-08	0.0248	0.0752	0.0752	0.1202	0.0747	0.0747	0.2043
HC_Aromatic_C36+	8.99E-18	0.0432	0.0049	0.0049	0.1482	0.0032	0.0032	0.0140
HC_Aromatic_Dehy_C10-C14	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C15-C19	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C20-C35	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C36+	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_AromaticN_C7	1.38E-05	2.11E-06	8.46E-05	8.46E-05	9.79E-06	8.56E-05	8.56E-05	2.70E-05
HC_AromaticN_C8	7.35E-06	2.80E-06	9.40E-05	9.40E-05	1.30E-05	9.50E-05	9.50E-05	3.60E-05
HC_AromaticN_C9	3.93E-06	3.68E-06	1.05E-04	1.05E-04	1.72E-05	1.06E-04	1.06E-04	4.77E-05
HC_AromaticN_C10-C14	6.88E-07	7.56E-06	1.38E-04	1.38E-04	3.55E-05	1.39E-04	1.39E-04	9.77E-05
HC_AromaticN_C15-C19	3.17E-08	2.31E-05	2.06E-04	2.06E-04	1.10E-04	2.08E-04	2.08E-04	2.76E-04
HC_AromaticN_C20-C35	9.75E-11	1.29E-04	3.54E-04	3.54E-04	6.24E-04	3.51E-04	3.51E-04	9.91E-04
HC_AromaticN_C36+	7.27E-20	0.0135	0.0010	0.0010	0.0394	5.44E-04	5.44E-04	0.0024
HC_Aromatics_C7	1.52E-04	2.32E-05	9.29E-04	9.29E-04	1.07E-04	9.39E-04	9.39E-04	2.97E-04
HC_Aromatics_C8	8.06E-05	3.06E-05	0.0010	0.0010	1.42E-04	0.0010	0.0010	3.95E-04
HC_Aromatics_C9	4.30E-05	4.03E-05	0.0011	0.0011	1.88E-04	0.0012	0.0012	5.21E-04
HC_Aromatics_C10-C14	7.47E-06	8.22E-05	0.0015	0.0015	3.85E-04	0.0015	0.0015	0.0011
HC_Aromatics_C15-C19	3.40E-07	2.48E-04	0.0022	0.0022	0.0012	0.0022	0.0022	0.0030
HC_Aromatics_C20-C35	1.02E-09	0.0014	0.0037	0.0037	0.0065	0.0037	0.0037	0.0104
HC_Aromatics_C36+	6.83E-19	0.1267	0.0095	0.0095	0.3702	0.0051	0.0051	0.0225

Components	S17	S18	S19	S20	S21	S22	S23	S24
T (°C)	380.0	240.0	70.0	70.0	70.0	70.0	70.0	72.7
P (kPa)	16800.00	16750.00	16700.00	16700.00	16700.00	16700.00	16700.00	17100.00
Vapour Fraction (-)	1.00	0.9361	0.8660	0.00	0.00	1.00	1.00	1.00
Mole Flow (k·mol/h)	30869.00	30869.29	30869.00	144.83	3991.25	26732.92	18713.04	18713.04
Mass Flow (kg/h)	645650.3	645646.0	645650.3	2608.63	442967.2	200074.6	140052.2	140052.2
Mass Fraction (-)								
HYDROGEN	0.0642	0.0642	0.0642	1.77E-04	0.0017	0.2033	0.2033	0.2033
METHANE	0.1149	0.1149	0.1149	1.06E-06	0.0105	0.3475	0.3475	0.3475
ETHANE	0.00	0.00	0.00	0.00E+00	0.00	0.00	0.00	0.00
PROPANE	0.0454	0.0454	0.0454	1.80E-10	0.0199	0.1023	0.1023	0.1023
n-BUTANE	0.0464	0.0464	0.0464	5.03E-12	0.0347	0.0732	0.0732	0.0732
CARBON MONOXIDE	0.00	0.00	0.00	0.00E+00	0.00	0.00	0.00	0.00
CARBON DIOXIDE	0.00	0.00	0.00	0.00E+00	0.00	0.00	0.00	0.00
NITROGEN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
WATER	0.0073	0.0073	0.0073	0.9505	5.19E-04	0.0101	0.0101	0.0101
AMMONIA	0.0041	0.0041	0.0041	0.0416	0.0016	0.0092	0.0092	0.0092
HYDROGEN SULFIDE	0.0765	0.0765	0.0765	0.0077	0.0229	0.1961	0.1961	0.1961
IRON (II) SULFATE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SULFUR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-HEXANE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Paraffin_C6	0.0494	0.0494	0.0494	2.92E-15	0.0587	0.0295	0.0295	0.0295
HC_Paraffin_C7	0.0455	0.0455	0.0455	6.69E-18	0.0598	0.0145	0.0145	0.0145
HC_Paraffin_C8	0.0431	0.0431	0.0431	1.27E-20	0.0596	0.0070	0.0070	0.0070
HC_Paraffin_C9	0.0414	0.0414	0.0414	1.82E-23	0.0589	0.0034	0.0034	0.0034
HC_Paraffin_C10-C14	0.1076	0.1076	0.1076	1.29E-31	0.1563	0.0012	0.0012	0.0012
HC_Paraffin_C15-C19	0.1485	0.1485	0.1485	3.94E-44	0.2164	5.06E-05	5.06E-05	5.06E-05
HC_Paraffin_C20-C35	0.1023	0.1023	0.1023	1.37E-44	0.1491	7.10E-08	7.10E-08	7.10E-08
HC_Paraffin_C36+	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_C6	0.0034	0.0034	0.0034	1.25E-09	0.0045	0.0012	0.0012	0.0012
HC_Aromatic_C7	0.0036	0.0036	0.0036	4.50E-11	0.0050	6.26E-04	6.26E-04	6.26E-04
HC_Aromatic_C8	0.0039	0.0039	0.0039	1.81E-12	0.0055	3.36E-04	3.36E-04	3.36E-04
HC_Aromatic_C9	0.0041	0.0041	0.0041	7.53E-14	0.0059	1.79E-04	1.79E-04	1.79E-04
HC_Aromatic_C10-C14	0.0137	0.0137	0.0137	3.89E-17	0.0200	8.92E-05	8.92E-05	8.92E-05
HC_Aromatic_C15-C19	0.0275	0.0275	0.0275	2.88E-23	0.0400	6.70E-06	6.70E-06	6.70E-06
HC_Aromatic_C20-C35	0.0370	0.0370	0.0370	5.30E-34	0.0540	2.97E-08	2.97E-08	2.97E-08
HC_Aromatic_C36+	5.26E-05	5.26E-05	5.26E-05	7.08E-48	7.66E-05	1.16E-17	1.16E-17	1.16E-17
HC_Aromatic_Dehy_C10-C14	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C15-C19	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C20-C35	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C36+	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_AromaticN_C7	1.03E-04	1.03E-04	1.03E-04	1.20E-12	1.41E-04	1.78E-05	1.78E-05	1.78E-05
HC_AromaticN_C8	1.12E-04	1.12E-04	1.12E-04	9.07E-14	1.59E-04	9.46E-06	9.46E-06	9.46E-06
HC_AromaticN_C9	1.23E-04	1.23E-04	1.23E-04	7.31E-15	1.76E-04	5.06E-06	5.06E-06	5.06E-06
HC_AromaticN_C10-C14	1.51E-04	1.51E-04	1.51E-04	8.93E-18	2.20E-04	8.85E-07	8.85E-07	8.85E-07
HC_AromaticN_C15-C19	1.88E-04	1.88E-04	1.88E-04	1.29E-22	2.74E-04	4.08E-08	4.08E-08	4.08E-08
HC_AromaticN_C20-C35	1.65E-04	1.65E-04	1.65E-04	1.41E-30	2.41E-04	1.25E-10	1.25E-10	1.25E-10
HC_AromaticN_C36+	5.33E-06	5.33E-06	5.33E-06	1.44E-48	7.77E-06	9.36E-20	9.36E-20	9.36E-20
HC_Aromatics_C7	0.0011	0.0011	0.0011	1.32E-11	0.0016	1.95E-04	1.95E-04	1.95E-04
HC_Aromatics_C8	0.0012	0.0012	0.0012	9.94E-13	0.0017	1.04E-04	1.04E-04	1.04E-04
HC_Aromatics_C9	0.0013	0.0013	0.0013	7.99E-14	0.0019	5.53E-05	5.53E-05	5.53E-05
HC_Aromatics_C10-C14	0.0016	0.0016	0.0016	9.70E-17	0.0024	9.61E-06	9.61E-06	9.61E-06
HC_Aromatics_C15-C19	0.0020	0.0020	0.0020	1.38E-21	0.0029	4.37E-07	4.37E-07	4.37E-07
HC_Aromatics_C20-C35	0.0017	0.0017	0.0017	1.48E-29	0.0025	1.31E-09	1.31E-09	1.31E-09
HC_Aromatics_C36+	5.01E-05	5.01E-05	5.01E-05	1.35E-47	7.30E-05	8.79E-19	8.79E-19	8.79E-19

Components	S25	S26	S27	S28	S29	S30	S31	S32
T (°C)	366.0	70.0	10.1	119.0	50.0	75.0	40.0	36.6
P (kPa)	17000.00	16700.00	5500.00	5500.00	5500.00	25000.00	5400.00	2600.00
Vapour Fraction (-)	1.00	1.00	0.9908	1.0000	0.9969	0.00E+00	1.0000	1.0000
Mole Flow (k·mol/h)	18713.08	8019.88	8019.88	9620.80	9310.50	726.21	8584.29	8584.29
Mass Flow (kg/h)	140052.3	1	60022.37	60022.37	93067.01	93067.00	22732.04	70334.97
Mass Fraction (-)								
HYDROGEN	0.2033	0.2033	0.2033	0.1450	0.1271	5.00E-04	0.1681	0.1681
METHANE	0.3475	0.3475	0.3475	0.2864	0.3911	0.0017	0.5169	0.5169
ETHANE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
PROPANE	0.1023	0.1023	0.1023	0.1306	0.1306	6.95E-04	0.1725	0.1725
n-BUTANE	0.0732	0.0732	0.0732	0.1060	0.1060	5.67E-04	0.1401	0.1401
CARBON MONOXIDE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CARBON DIOXIDE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NITROGEN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
WATER	0.0101	0.0101	0.0101	0.0105	0.0105	0.0357	0.0024	0.0024
AMMONIA	0.0092	0.0092	0.0092	0.0116	0.0116	0.0475	0.00	0.00
HYDROGEN SULFIDE	0.1961	0.1961	0.1961	0.2229	0.2231	0.9134	0.00	0.00
IRON (II) SULFATE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SULFUR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-HEXANE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Paraffin_C6	0.0295	0.0295	0.0295	0.0449	0.00	0.00	0.00	0.00
HC_Paraffin_C7	0.0145	0.0145	0.0145	0.0213	0.00	0.00	0.00	0.00
HC_Paraffin_C8	0.0070	0.0070	0.0070	0.0098	0.00	0.00	0.00	0.00
HC_Paraffin_C9	0.0034	0.0034	0.0034	0.0045	0.00	0.00	0.00	0.00
HC_Paraffin_C10-C14	0.0012	0.0012	0.0012	0.0015	0.00	0.00	0.00	0.00
HC_Paraffin_C15-C19	5.06E-05	5.06E-05	5.06E-05	1.78E-04	0.00	0.00	0.00	0.00
HC_Paraffin_C20-C35	7.10E-08	7.10E-08	7.10E-08	5.73E-05	0.00	0.00	0.00	0.00
HC_Paraffin_C36+	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_C6	0.0012	0.0012	0.0012	0.0020	0.00	0.00	0.00	0.00
HC_Aromatic_C7	6.26E-04	6.26E-04	6.26E-04	0.0010	0.00	0.00	0.00	0.00
HC_Aromatic_C8	3.36E-04	3.36E-04	3.36E-04	5.45E-04	0.00	0.00	0.00	0.00
HC_Aromatic_C9	1.79E-04	1.79E-04	1.79E-04	2.85E-04	0.00	0.00	0.00	0.00
HC_Aromatic_C10-C14	8.92E-05	8.92E-05	8.92E-05	1.47E-04	0.00	0.00	0.00	0.00
HC_Aromatic_C15-C19	6.70E-06	6.70E-06	6.70E-06	3.53E-05	0.00	0.00	0.00	0.00
HC_Aromatic_C20-C35	2.97E-08	2.97E-08	2.97E-08	2.90E-05	0.00	0.00	0.00	0.00
HC_Aromatic_C36+	1.16E-17	1.16E-17	1.16E-17	1.46E-07	1.46E-07	0.00	1.93E-07	1.93E-07
HC_Aromatic_Dehy_C10-C14	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C15-C19	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C20-C35	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C36+	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_AromaticN_C7	1.78E-05	1.78E-05	1.78E-05	2.96E-05	0.00	0.00	0.00	0.00
HC_AromaticN_C8	9.46E-06	9.46E-06	9.46E-06	1.54E-05	0.00	0.00	0.00	0.00
HC_AromaticN_C9	5.06E-06	5.06E-06	5.06E-06	8.15E-06	0.00	0.00	0	0.00
HC_AromaticN_C10-C14	8.85E-07	8.85E-07	8.85E-07	1.51E-06	1.51E-06	0.00	2.00E-06	2.00E-06
HC_AromaticN_C15-C19	4.08E-08	4.08E-08	4.08E-08	2.39E-07	2.39E-07	0.00	3.16E-07	3.16E-07
HC_AromaticN_C20-C35	1.25E-10	1.25E-10	1.25E-10	1.37E-07	1.37E-07	0.00	1.82E-07	1.82E-07
HC_AromaticN_C36+	9.36E-20	9.36E-20	9.36E-20	1.99E-08	1.99E-08	0.00	2.64E-08	2.64E-08
HC_Aromatics_C7	1.95E-04	1.95E-04	1.95E-04	3.25E-04	0.00	0.00	0.00	0.00
HC_Aromatics_C8	1.04E-04	1.04E-04	1.04E-04	1.69E-04	0.00	0.00	0.00	0.00
HC_Aromatics_C9	5.53E-05	5.53E-05	5.53E-05	8.91E-05	0.00	0.00	0.00	0.00
HC_Aromatics_C10-C14	9.61E-06	9.61E-06	9.61E-06	1.65E-05	0.00	0.00	0.00	0.00
HC_Aromatics_C15-C19	4.37E-07	4.37E-07	4.37E-07	2.57E-06	2.57E-06	0.00	3.40E-06	3.40E-06
HC_Aromatics_C20-C35	1.31E-09	1.31E-09	1.31E-09	1.44E-06	1.44E-06	0.00	1.90E-06	1.90E-06
HC_Aromatics_C36+	8.79E-19	8.79E-19	8.79E-19	1.87E-07	1.87E-07	0.00	2.48E-07	2.48E-07

Components	S33	S34	S35	S36	S37	S38	S39	S40
T (°C)	300.0	200.0	80.0	80.0	342.0	453.3	455.3	350.0
P (kPa)	2600.00	2600.00	500.00	2100.00	17000.00	16900.00	7900.00	7900.00
Vapour Fraction (-)	1.00	1.00	1.0000	1.00	1.00	0.0011	0.1857	0.1867
Mole Flow (k·mol/h)	7715.67	2162.32	10551.84	19929.99	19929.99	81.13	81.13	81.13
Mass Flow (kg/h)	139000	39845.00	207918.1	40176.46	40176.46	19339.80	19339.8	19339.80
Mass Fraction (-)								
HYDROGEN	0.00	0.00	0.0341	1.00	1.00	0.0007	0.0007	0.0007
METHANE	0.00	0.7835	0.0201	0.00	0.00	0.0016	0.0016	0.0016
ETHANE	0.00	0.0816	1.81E-06	0.00	0.00	0.00	0.00	0.00
PROPANE	0.00	0.0718	1.51E-11	0.00	0.00	0.0010	0.0010	0.0010
n-BUTANE	0.00	0.0631	5.94E-12	0.00	0.00	0.0013	0.0013	0.0013
CARBON MONOXIDE	0.00	0.00	0.6938	0.00	0.00	0.00	0.00	0.00
CARBON DIOXIDE	0.00	0.00	0.1884	0.00	0.00	0.00	0.00	0.00
NITROGEN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
WATER	1.00	0.00	0.0637	0.00	0.00	1.41E-04	1.41E-04	1.41E-04
AMMONIA	0.00	0.00	0.00	0.00	0.00	7.05E-05	7.05E-05	7.05E-05
HYDROGEN SULFIDE	0.00	0.00	0.00	0.00	0.00	0.0016	0.0016	0.0016
IRON (II) SULFATE	0.00	0.00	0.00	0.00	0.00	0.34	0.34	0.34
SULFUR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-HEXANE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Paraffin_C6	0.00	0.00	0.00	0.00	0.00	0.0019	0.0019	0.0019
HC_Paraffin_C7	0.00	0.00	0.00	0.00	0.00	0.0020	0.0020	0.0020
HC_Paraffin_C8	0.00	0.00	0.00	0.00	0.00	0.0022	0.0022	0.0022
HC_Paraffin_C9	0.00	0.00	0.00	0.00	0.00	0.0025	0.0025	0.0025
HC_Paraffin_C10-C14	0.00	0.00	0.00	0.00	0.00	0.0101	0.0101	0.0101
HC_Paraffin_C15-C19	0.00	0.00	0.00	0.00	0.00	0.0317	0.0317	0.0317
HC_Paraffin_C20-C35	0.00	0.00	0.00	0.00	0.00	0.1260	0.1260	0.1260
HC_Paraffin_C36+	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_C6	0.00	0.00	0.00	0.00	0.00	1.70E-04	1.70E-04	1.70E-04
HC_Aromatic_C7	0.00	0.00	0.00	0.00	0.00	2.17E-04	2.17E-04	2.17E-04
HC_Aromatic_C8	0.00	0.00	0.00	0.00	0.00	2.80E-04	2.80E-04	2.80E-04
HC_Aromatic_C9	0.00	0.00	0.00	0.00	0.00	3.55E-04	3.55E-04	3.55E-04
HC_Aromatic_C10-C14	0.00	0.00	0.00	0.00	0.00	0.0019	0.0019	0.0019
HC_Aromatic_C15-C19	0.00	0.00	0.00	0.00	0.00	0.0092	0.0092	0.0092
HC_Aromatic_C20-C35	0.00	0.00	0.00	0.00	0.00	0.0749	0.0749	0.0749
HC_Aromatic_C36+	0.00	0.00	5.23E-14	0.00	0.00	0.0984	0.0984	0.0984
HC_Aromatic_Dehy_C10-C14	0.00	0.00	0.00E+00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C15-C19	0.00	0.00	0.00E+00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C20-C35	0.00	0.00	0.00E+00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C36+	0.00	0.00	0.00E+00	0.00	0.00	0.00	0.00	0.00
HC_AromaticN_C7	0.00	0.00	0.00E+00	0.00	0.00	6.15E-06	6.15E-06	6.15E-06
HC_AromaticN_C8	0.00	0.00	0.00E+00	0.00	0.00	8.15E-06	8.15E-06	8.15E-06
HC_AromaticN_C9	0.00	0.00	0.00E+00	0.00	0.00	1.08E-05	1.08E-05	1.08E-05
HC_AromaticN_C10-C14	0.00	0.00	6.78E-07	0.00	0.00	2.22E-05	2.22E-05	2.22E-05
HC_AromaticN_C15-C19	0.00	0.00	1.07E-07	0.00	0.00	6.85E-05	6.85E-05	6.85E-05
HC_AromaticN_C20-C35	0.00	0.00	5.92E-08	0.00	0.00	3.89E-04	3.89E-04	3.89E-04
HC_AromaticN_C36+	0.00	0.00	8.90E-09	0.00	0.00	0.0271	0.0271	0.0271
HC_Aromatics_C7	0.00	0.00	0.00	0.00	0.00	6.75E-05	6.75E-05	6.75E-05
HC_Aromatics_C8	0.00	0.00	0.00	0.00	0.00	8.93E-05	8.93E-05	8.93E-05
HC_Aromatics_C9	0.00	0.00	0.00	0.00	0.00	1.18E-04	1.18E-04	1.18E-04
HC_Aromatics_C10-C14	0.00	0.00	0.00	0.00	0.00	2.42E-04	2.42E-04	2.42E-04
HC_Aromatics_C15-C19	0.00	0.00	1.15E-06	0.00	0.00	0.0007	0.0007	0.0007
HC_Aromatics_C20-C35	0.00	0.00	6.20E-07	0.00	0.00	0.0041	0.0041	0.0041
HC_Aromatics_C36+	0.00	0.00	1.33E-14	0.00	0.00	0.2547	0.2547	0.2547

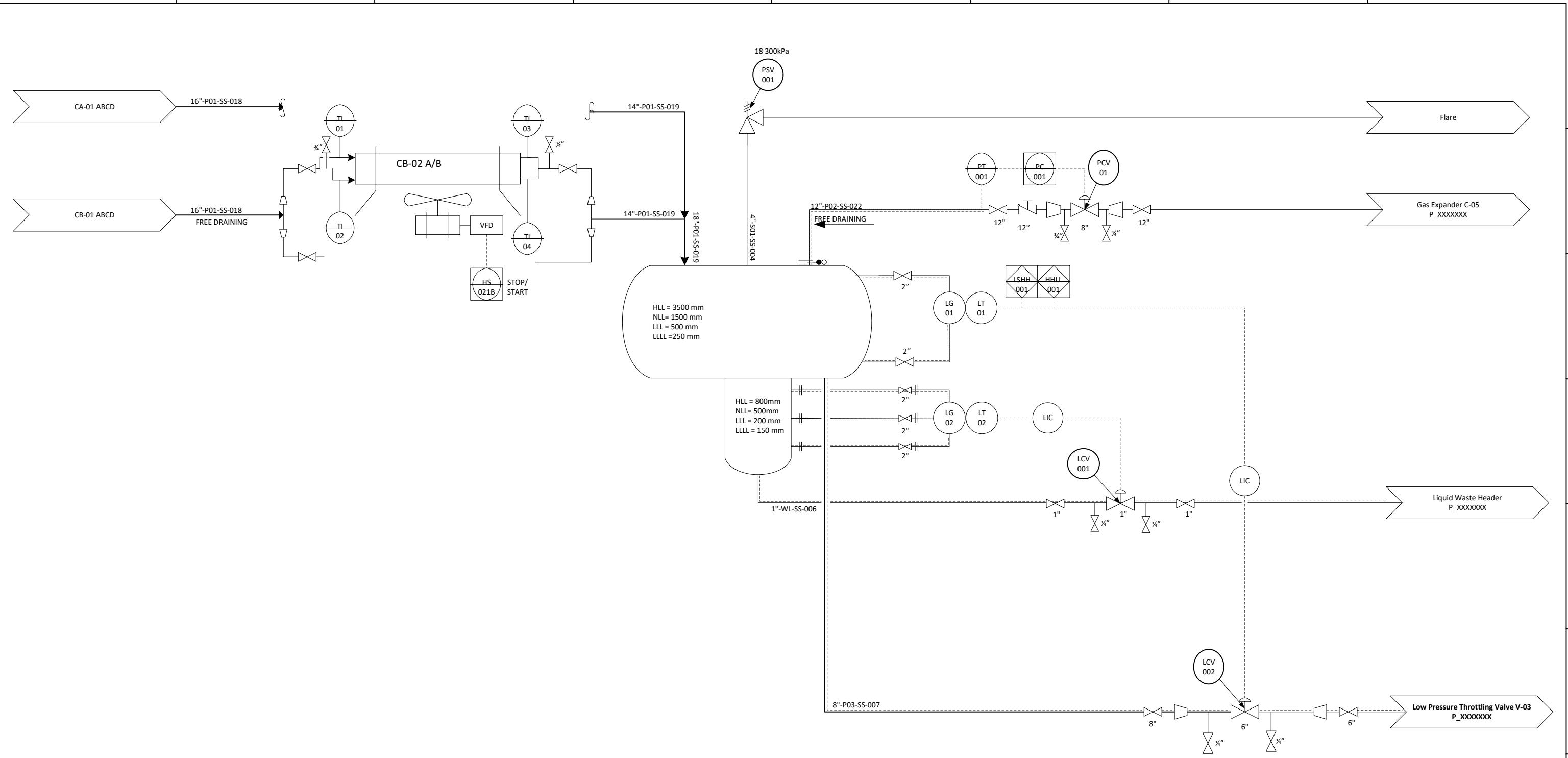
Components	S41	S42	S43	S44	S45	S46	S47	S48
T (°C)	350.0	350.0	350.0	100.0	197.2	319.0	80.0	80.0
P (kPa)	7800.00	7800.00	7800.00	100.00	100.00	100.00	90.00	90.00
Vapour Fraction (-)	0.1881	1.00	0.00	1.00	2.21E-08	1.0000	0.0310	1.00
Mole Flow (k·mol/h)	81.13	7.02	74.11	249.79	53.56	270.34	270.34	8.38
Mass Flow (kg/h)	19339.8	0	126.88	19212.92	4500.00	14424.41	9288.51	9288.51
Mass Fraction (-)								175.28
HYDROGEN	0.0007	0.0719	1.90E-04	0.00	2.23E-27	3.93E-04	3.93E-04	0.0208
METHANE	0.0016	0.1465	6.46E-04	0.00	7.65E-25	0.0013	0.0013	0.0704
ETHANE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
PROPANE	0.0010	0.0631	5.78E-04	0.00	1.97E-22	0.0012	0.0012	0.0599
n-BUTANE	0.0013	0.0644	0.0008	0.00	4.27E-21	0.0017	0.0017	0.0787
CARBON MONOXIDE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CARBON DIOXIDE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NITROGEN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
WATER	1.41E-04	0.0099	7.63E-05	1.00	7.58E-05	0.4845	0.4845	0.4522
AMMONIA	7.05E-05	0.0053	3.61E-05	0.00	5.66E-24	7.47E-05	7.47E-05	0.0011
HYDROGEN SULFIDE	0.0016	0.1075	0.0009	0.00	2.87E-22	0.0019	0.0019	0.0944
IRON (II) SULFATE	0.34	0.00	0.35	0.00	4.61E-01	0.00	0.00	0.00
SULFUR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-HEXANE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Paraffin_C6	0.0019	0.0650	0.0015	0.00	5.97E-19	0.0030	0.0030	0.0847
HC_Paraffin_C7	0.0020	0.0570	0.0017	0.00	5.28E-18	0.0034	0.0034	0.0576
HC_Paraffin_C8	0.0022	0.0509	0.0019	0.00	4.88E-17	0.0039	0.0039	0.0343
HC_Paraffin_C9	0.0025	0.0459	0.0022	0.00	4.56E-16	0.0046	0.0046	0.0187
HC_Paraffin_C10-C14	0.0101	0.0973	0.0095	0.00	6.78E-13	0.0197	0.0197	0.0080
HC_Paraffin_C15-C19	0.0317	0.0933	0.0313	0.00	5.07E-08	0.0647	0.0647	4.19E-04
HC_Paraffin_C20-C35	0.1260	0.0420	0.1266	0.00	0.0239	0.2248	0.2248	9.89E-07
HC_Paraffin_C36+	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_C6	1.70E-04	0.0045	1.42E-04	0.00	4.58E-19	2.93E-04	2.93E-04	0.0058
HC_Aromatic_C7	2.17E-04	0.0045	1.89E-04	0.00	5.43E-18	3.92E-04	3.92E-04	0.0042
HC_Aromatic_C8	2.80E-04	0.0045	2.52E-04	0.00	5.74E-17	5.22E-04	5.22E-04	0.0028
HC_Aromatic_C9	3.55E-04	0.0045	3.28E-04	0.00	5.47E-16	6.79E-04	6.79E-04	0.0017
HC_Aromatic_C10-C14	0.0019	0.0125	0.0019	0.00	6.11E-13	0.0038	0.0038	0.0011
HC_Aromatic_C15-C19	0.0092	0.0191	0.0091	0.00	1.56E-08	0.0189	0.0189	1.39E-04
HC_Aromatic_C20-C35	0.0749	0.0213	0.0753	0.00	0.0054	0.1473	0.1473	1.61E-06
HC_Aromatic_C36+	0.0984	1.07E-04	0.0990	0.00	0.1315	5.55E-04	5.55E-04	1.78E-16
HC_Aromatic_Dehy_C10-C14	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C15-C19	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C20-C35	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C36+	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_AromaticN_C7	6.15E-06	1.26E-04	5.35E-06	0.00	1.53E-19	1.11E-05	1.11E-05	1.19E-04
HC_AromaticN_C8	8.15E-06	1.30E-04	7.35E-06	0.00	1.80E-18	1.52E-05	1.52E-05	7.91E-05
HC_AromaticN_C9	1.08E-05	1.33E-04	9.95E-06	0.00	1.93E-17	2.06E-05	2.06E-05	4.96E-05
HC_AromaticN_C10-C14	2.22E-05	1.37E-04	2.15E-05	0.00	9.76E-15	4.44E-05	4.44E-05	1.22E-05
HC_AromaticN_C15-C19	6.85E-05	1.31E-04	6.81E-05	0.00	1.66E-10	1.41E-04	1.41E-04	9.96E-07
HC_AromaticN_C20-C35	3.89E-04	1.01E-04	3.91E-04	0.00	3.13E-05	7.61E-04	7.61E-04	9.19E-09
HC_AromaticN_C36+	0.0271	1.46E-05	0.0273	0.00	0.0363	7.43E-05	7.43E-05	2.21E-18
HC_Aromatics_C7	6.75E-05	0.0014	5.88E-05	0.00	1.67E-18	1.22E-04	1.22E-04	0.0013
HC_Aromatics_C8	8.93E-05	0.0014	8.05E-05	0.00	1.97E-17	1.67E-04	1.67E-04	8.67E-04
HC_Aromatics_C9	1.18E-04	0.0015	1.09E-04	0.00	2.11E-16	2.25E-04	2.25E-04	5.42E-04
HC_Aromatics_C10-C14	2.42E-04	0.0015	2.33E-04	0.00	1.06E-13	4.83E-04	4.83E-04	1.33E-04
HC_Aromatics_C15-C19	0.0007	0.0014	0.0007	0.00	1.79E-09	0.0015	0.0015	1.07E-05
HC_Aromatics_C20-C35	0.0041	0.0011	0.0041	0.00	3.27E-04	0.0080	0.0080	9.61E-08
HC_Aromatics_C36+	0.2547	1.37E-04	0.2564	0.00	0.3410	6.98E-04	6.98E-04	2.08E-17

Components	S49	S50	S51	S52	S53	S54	S55	S56
T (°C)	80.0	80.1	80.0	75.9	172.8	173.0	80.0	80.0
P (kPa)	90.00	490.00	90.00	90.00	16700.00	2000.00	2000.00	2000.00
Vapour Fraction (-)	0.00	0.00	0.00	0.0048	0.0018	0.3498	0.2481	1.00
Mole Flow (k·mol/h)	16.59	16.59	245.37	390.19	5077.00	5077.00	5077.00	1259.58
Mass Flow (kg/h)	4692.76	4692.76	4420.47	7029.10	630376.6	630376.6	630376.	20832.69
Mass Fraction (-)								
HYDROGEN	1.04E-06	1.04E-06	2.63E-07	6.59E-05	0.0020	0.0020	0.0020	0.0585
METHANE	1.58E-05	1.58E-05	4.15E-09	3.97E-07	0.0094	0.0094	0.0094	0.2337
ETHANE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
PROPANE	1.29E-04	1.29E-04	2.92E-12	6.87E-11	0.0153	0.0153	0.0153	0.1664
n-BUTANE	4.98E-04	4.98E-04	1.82E-13	1.98E-12	0.0260	0.0260	0.0260	0.1341
CARBON MONOXIDE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CARBON DIOXIDE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NITROGEN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
WATER	3.15E-04	3.15E-04	0.9998	0.9815	5.24E-04	5.24E-04	5.24E-04	0.0101
AMMONIA	1.86E-06	1.86E-06	1.11E-04	0.0155	0.0012	0.0012	0.0012	0.0151
HYDROGEN SULFIDE	1.06E-04	1.06E-04	7.80E-05	0.0029	0.0180	0.0180	0.0180	0.2771
IRON (II) SULFATE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SULFUR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-HEXANE	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Paraffin_C6	0.0029	0.0029	3.46E-16	1.30E-15	0.0438	0.0438	0.0438	0.0554
HC_Paraffin_C7	0.0046	0.0046	1.32E-18	3.31E-18	0.0448	0.0448	0.0448	0.0255
HC_Paraffin_C8	0.0065	0.0065	3.72E-21	7.06E-21	0.0450	0.0450	0.0450	0.0114
HC_Paraffin_C9	0.0084	0.0084	7.28E-24	1.13E-23	0.0449	0.0449	0.0449	0.0051
HC_Paraffin_C10-C14	0.0387	0.0387	1.06E-31	1.15E-31	0.1242	0.1242	0.1242	0.0015
HC_Paraffin_C15-C19	0.1280	0.1280	6.02E-44	5.25E-44	0.1965	0.1965	0.1965	4.50E-05
HC_Paraffin_C20-C35	0.4449	0.4449	1.05E-43	7.08E-44	0.2310	0.2310	0.2310	4.76E-08
HC_Paraffin_C36+	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_C6	3.62E-04	3.62E-04	1.77E-10	5.75E-10	0.0034	0.0034	0.0034	0.0026
HC_Aromatic_C7	6.18E-04	6.18E-04	9.79E-12	2.29E-11	0.0038	0.0038	0.0038	0.0013
HC_Aromatic_C8	9.29E-04	9.29E-04	5.40E-13	1.01E-12	0.0042	0.0042	0.0042	6.87E-04
HC_Aromatic_C9	0.0013	0.0013	2.90E-14	4.62E-14	0.0046	0.0046	0.0046	3.51E-04
HC_Aromatic_C10-C14	0.0076	0.0076	2.68E-17	3.13E-17	0.0166	0.0166	0.0166	1.60E-04
HC_Aromatic_C15-C19	0.0374	0.0374	4.99E-23	4.21E-23	0.0395	0.0395	0.0395	1.13E-05
HC_Aromatic_C20-C35	0.2915	0.2915	4.90E-33	3.28E-33	0.0987	0.0987	0.0987	6.17E-08
HC_Aromatic_C36+	0.0011	0.0011	2.58E-46	1.65E-46	0.0042	0.0042	0.0042	2.33E-16
HC_Aromatic_Dehy_C10-C14	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C15-C19	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C20-C35	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C36+	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_AromaticN_C7	1.75E-05	1.75E-05	2.61E-13	6.09E-13	1.07E-04	1.07E-04	1.07E-04	3.79E-05
HC_AromaticN_C8	2.71E-05	2.71E-05	2.70E-14	5.07E-14	1.22E-04	1.22E-04	1.22E-04	1.95E-05
HC_AromaticN_C9	3.89E-05	3.89E-05	2.80E-15	4.47E-15	1.38E-04	1.38E-04	1.38E-04	1.01E-05
HC_AromaticN_C10-C14	8.75E-05	8.75E-05	6.05E-18	7.12E-18	1.84E-04	1.84E-04	1.84E-04	1.66E-06
HC_AromaticN_C15-C19	2.79E-04	2.79E-04	2.20E-22	1.86E-22	2.74E-04	2.74E-04	2.74E-04	7.64E-08
HC_AromaticN_C20-C35	0.0015	0.0015	1.22E-29	8.19E-30	4.64E-04	4.64E-04	4.64E-04	3.29E-10
HC_AromaticN_C36+	1.47E-04	1.47E-04	6.72E-47	4.28E-47	7.19E-04	7.19E-04	7.19E-04	4.90E-18
HC_Aromatics_C7	1.92E-04	1.92E-04	2.87E-12	6.69E-12	0.0012	0.0012	0.0012	4.16E-04
HC_Aromatics_C8	2.97E-04	2.97E-04	2.96E-13	5.55E-13	0.0013	0.0013	0.0013	2.14E-04
HC_Aromatics_C9	4.25E-04	4.25E-04	3.06E-14	4.89E-14	0.0015	0.0015	0.0015	1.10E-04
HC_Aromatics_C10-C14	9.50E-04	9.50E-04	6.57E-17	7.73E-17	0.0020	0.0020	0.0020	1.81E-05
HC_Aromatics_C15-C19	0.0030	0.0030	2.36E-21	2.00E-21	0.0029	0.0029	0.0029	8.20E-07
HC_Aromatics_C20-C35	0.0158	0.0158	1.28E-28	8.57E-29	0.0049	0.0049	0.0049	3.45E-09
HC_Aromatics_C36+	0.0014	0.0014	6.31E-46	4.02E-46	0.0068	0.0068	0.0068	4.60E-17

Components	S57	S58	S59	S60	S61	S62	S63
T (°C)	80.0	78.2	78.2	78.2	78.2	76.0	460.0
P (kPa)	2000.00	400.00	400.00	400.00	400.00	90.00	5500.00
Vapour Fraction (-)	0.00	0.0854	0.00	0.00	1.00	1.00	1.00
Mole Flow (k·mol/h)	3817.42	3817.42	3491.47	3508.07	325.95	1600.93	1600.93
Mass Flow (kg/h)	609543.9	609543.9	597634.1	602326.9	11909.79	33044.64	33044.6
Mass Fraction (-)							
HYDROGEN	1.04E-04	1.04E-04	3.78E-06	3.75E-06	0.0051	0.0391	0.0391
METHANE	0.0017	0.0017	2.39E-04	2.37E-04	0.0754	0.1754	0.1754
ETHANE	0.00	0.00	0.00	0.00	0.00	0.00	0.00
PROPANE	0.0101	0.0101	0.0061	0.0060	0.2119	0.1819	0.1819
n-BUTANE	0.0223	0.0223	0.0183	0.0181	0.2234	0.1657	0.1657
CARBON MONOXIDE	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CARBON DIOXIDE	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NITROGEN	0.00	0.00	0.00	0.00	0.00	0.00	0.00
WATER	1.96E-04	1.96E-04	6.13E-05	6.33E-05	0.0069	0.0113	0.0113
AMMONIA	7.43E-04	7.43E-04	4.04E-04	4.00E-04	0.0178	0.0160	0.0160
HYDROGEN SULFIDE	0.0091	0.0091	0.0040	0.0040	0.2662	0.2716	0.2716
IRON (II) SULFATE	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SULFUR	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-HEXANE	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Paraffin_C6	0.0434	0.0434	0.0422	0.0419	0.1031	0.0728	0.0728
HC_Paraffin_C7	0.0454	0.0454	0.0454	0.0451	0.0470	0.0335	0.0335
HC_Paraffin_C8	0.0461	0.0461	0.0466	0.0463	0.0205	0.0150	0.0150
HC_Paraffin_C9	0.0462	0.0462	0.0470	0.0467	0.0089	0.0067	0.0067
HC_Paraffin_C10-C14	0.1284	0.1284	0.1309	0.1302	0.0023	0.0022	0.0022
HC_Paraffin_C15-C19	0.2032	0.2032	0.2073	0.2067	5.87E-05	4.10E-04	4.10E-04
HC_Paraffin_C20-C35	0.2388	0.2388	0.2436	0.2452	4.38E-08	1.61E-04	1.61E-04
HC_Paraffin_C36+	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_C6	0.0034	0.0034	0.0034	0.0033	0.0051	0.0035	0.0035
HC_Aromatic_C7	0.0039	0.0039	0.0039	0.0039	0.0025	0.0018	0.0018
HC_Aromatic_C8	0.0044	0.0044	0.0044	0.0044	0.0013	9.26E-04	9.26E-04
HC_Aromatic_C9	0.0048	0.0048	0.0049	0.0048	6.38E-04	4.78E-04	4.78E-04
HC_Aromatic_C10-C14	0.0171	0.0171	0.0175	0.0174	2.70E-04	2.53E-04	2.53E-04
HC_Aromatic_C15-C19	0.0408	0.0408	0.0416	0.0416	1.67E-05	8.72E-05	8.72E-05
HC_Aromatic_C20-C35	0.1021	0.1021	0.1041	0.1056	7.29E-08	8.17E-05	8.17E-05
HC_Aromatic_C36+	0.0044	0.0044	0.0044	0.0044	1.49E-16	4.10E-07	4.10E-07
HC_Aromatic_Dehy_C10-C14	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C15-C19	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C20-C35	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C36+	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HC_AromaticN_C7	1.10E-04	1.10E-04	1.11E-04	1.10E-04	7.22E-05	5.11E-05	5.11E-05
HC_AromaticN_C8	1.26E-04	1.26E-04	1.28E-04	1.27E-04	3.64E-05	2.63E-05	2.63E-05
HC_AromaticN_C9	1.42E-04	1.42E-04	1.45E-04	1.44E-04	1.84E-05	1.38E-05	1.38E-05
HC_AromaticN_C10-C14	1.90E-04	1.90E-04	1.94E-04	1.93E-04	2.82E-06	2.66E-06	2.66E-06
HC_AromaticN_C15-C19	2.84E-04	2.84E-04	2.89E-04	2.89E-04	1.15E-07	6.00E-07	6.00E-07
HC_AromaticN_C20-C35	4.80E-04	4.80E-04	4.89E-04	4.97E-04	4.04E-10	3.87E-07	3.87E-07
HC_AromaticN_C36+	7.43E-04	7.43E-04	7.58E-04	7.53E-04	3.25E-18	5.61E-08	5.61E-08
HC_Aromatics_C7	0.0012	0.0012	0.0012	0.0012	7.93E-04	5.61E-04	5.61E-04
HC_Aromatics_C8	0.0014	0.0014	0.0014	0.0014	3.98E-04	2.88E-04	2.88E-04
HC_Aromatics_C9	0.0016	0.0016	0.0016	0.0016	2.01E-04	1.50E-04	1.50E-04
HC_Aromatics_C10-C14	0.0021	0.0021	0.0021	0.0021	3.06E-05	2.89E-05	2.89E-05
HC_Aromatics_C15-C19	0.0030	0.0030	0.0031	0.0031	1.23E-06	6.43E-06	6.43E-06
HC_Aromatics_C20-C35	0.0050	0.0050	0.0051	0.0052	4.22E-09	4.05E-06	4.05E-06
HC_Aromatics_C36+	0.0070	0.0070	0.0071	0.0071	3.06E-17	5.27E-07	5.27E-07

Components	CT-S01	CT-S02	CT-S03	CP-S01
T (°C)	245.3	80.0	250.0	100.0
P (kPa)	5985.00	5885.00	17100.00	100.00
Vapour Fraction (-)	1.00	1.00	1.00	1.00
Mole Flow (k-mol/h)	19930.0	19929.99	19929.99	1407.48
Mass Flow (kg/h)	40176.5	40176.46	40176.46	23796.96
Mass Fraction (-)				
HYDROGEN	1.00	1.00	1.00	0.06
METHANE	0.00	0.00	0.00	0.00
ETHANE	0.00	0.00	0.00	0.00
PROPANE	0.00	0.00	0.00	0.00
n-BUTANE	0.00	0.00	0.00	0.00
CARBON MONOXIDE	0.00	0.00	0.00	0.00
CARBON DIOXIDE	0.00	0.00	0.00	0.01
NITROGEN	0.00	0.00	0.00	0.04
WATER	0.00	0.00	0.00	0.05
AMMONIA	0.00	0.00	0.00	0.00
HYDROGEN SULFIDE	0.00	0.00	0.00	0.00
IRON (II) SULFATE	0.00	0.00	0.00	0.00
SULFUR	0.00	0.00	0.00	0.85
n-HEXANE	0.00	0.00	0.00	0.00
HC_Paraffin_C6	0.00	0.00	0.00	0.00
HC_Paraffin_C7	0.00	0.00	0.00	0.00
HC_Paraffin_C8	0.00	0.00	0.00	0.00
HC_Paraffin_C9	0.00	0.00	0.00	0.00
HC_Paraffin_C10-C14	0.00	0.00	0.00	0.00
HC_Paraffin_C15-C19	0.00	0.00	0.00	0.00
HC_Paraffin_C20-C35	0.00	0.00	0.00	0.00
HC_Paraffin_C36+	0.00	0.00	0.00	0.00
HC_Aromatic_C6	0.00	0.00	0.00	0.00
HC_Aromatic_C7	0.00	0.00	0.00	0.00
HC_Aromatic_C8	0.00	0.00	0.00	0.00
HC_Aromatic_C9	0.00	0.00	0.00	0.00
HC_Aromatic_C10-C14	0.00	0.00	0.00	0.00
HC_Aromatic_C15-C19	0.00	0.00	0.00	0.00
HC_Aromatic_C20-C35	0.00	0.00	0.00	0.00
HC_Aromatic_C36+	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C10-C14	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C15-C19	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C20-C35	0.00	0.00	0.00	0.00
HC_Aromatic_Dehy_C36+	0.00	0.00	0.00	0.00
HC_AromaticN_C7	0.00	0.00	0.00	0.00
HC_AromaticN_C8	0.00	0.00	0.00	0.00
HC_AromaticN_C9	0.00	0.00	0.00	0.00
HC_AromaticN_C10-C14	0.00	0.00	0.00	0.00
HC_AromaticN_C15-C19	0.00	0.00	0.00	0.00
HC_AromaticN_C20-C35	0.00	0.00	0.00	0.00
HC_AromaticN_C36+	0.00	0.00	0.00	0.00
HC_Aromatics_C7	0.00	0.00	0.00	0.00
HC_Aromatics_C8	0.00	0.00	0.00	0.00
HC_Aromatics_C9	0.00	0.00	0.00	0.00
HC_Aromatics_C10-C14	0.00	0.00	0.00	0.00
HC_Aromatics_C15-C19	0.00	0.00	0.00	0.00
HC_Aromatics_C20-C35	0.00	0.00	0.00	0.00
HC_Aromatics_C36+	0.00	0.00	0.00	0.00

D.3 PROCESS & INSTRUMENTATION DIAGRAM



Reference Drawing Number	Reference Drawing Description	REVISIONS			
		REV	DESCRIPTION	DATE	APPROVED
		1			
		2			
P&ID, AND DRAWING STANDARD (2).VSDX					



Drawn By: Jose Te Eng Fo

Reviewed By: Yichun Zhang

SIZE	FSCM NO	DWG NO	REV
		TR-P-DWG-005	0
SCALE		SHEET	1 OF 1

D.3.1 P&ID Line Designation

Table D2. P&ID line designation table.

	Line Designation			Description			Mechanical				Operating Conditions	
#	Line #	Spec	Line Size	From	To	Service	Pipe Schedule	E.T	Pipe Material	Insulation	Temp (°C)	Pressure (kPa)
1	P-01	P	16"	Cooler CB-01	Cooler CB-02	Petro	120	Y	S.S	Cell Glass	270	16900
2	P-02	P	14"	From Cooler CB-02	To Mixpoint (Line 1 and 2)	Petro	100	Y	S.S	Cell Glass	70	16800
3	P-03	P	18"	From Mixpoint (Line 1 and 2)	To D-03	Petro	100	Y	S.S	Cell Glass	70	16800
4	P-04	P	12"	From D-03	Gas Expander EX-01	Petro	100	Y	S.S	Cell Glass	70	16800
5	P-05	P	8"	From D-03	Low Pressure Separator D-04	Petro	100	Y	S.S	Cell Glass	70	16800
6	W-01	W	1"	From D-03	Wastewater Drain	Water	40	Y	S.S	Cell Glass	70	16800

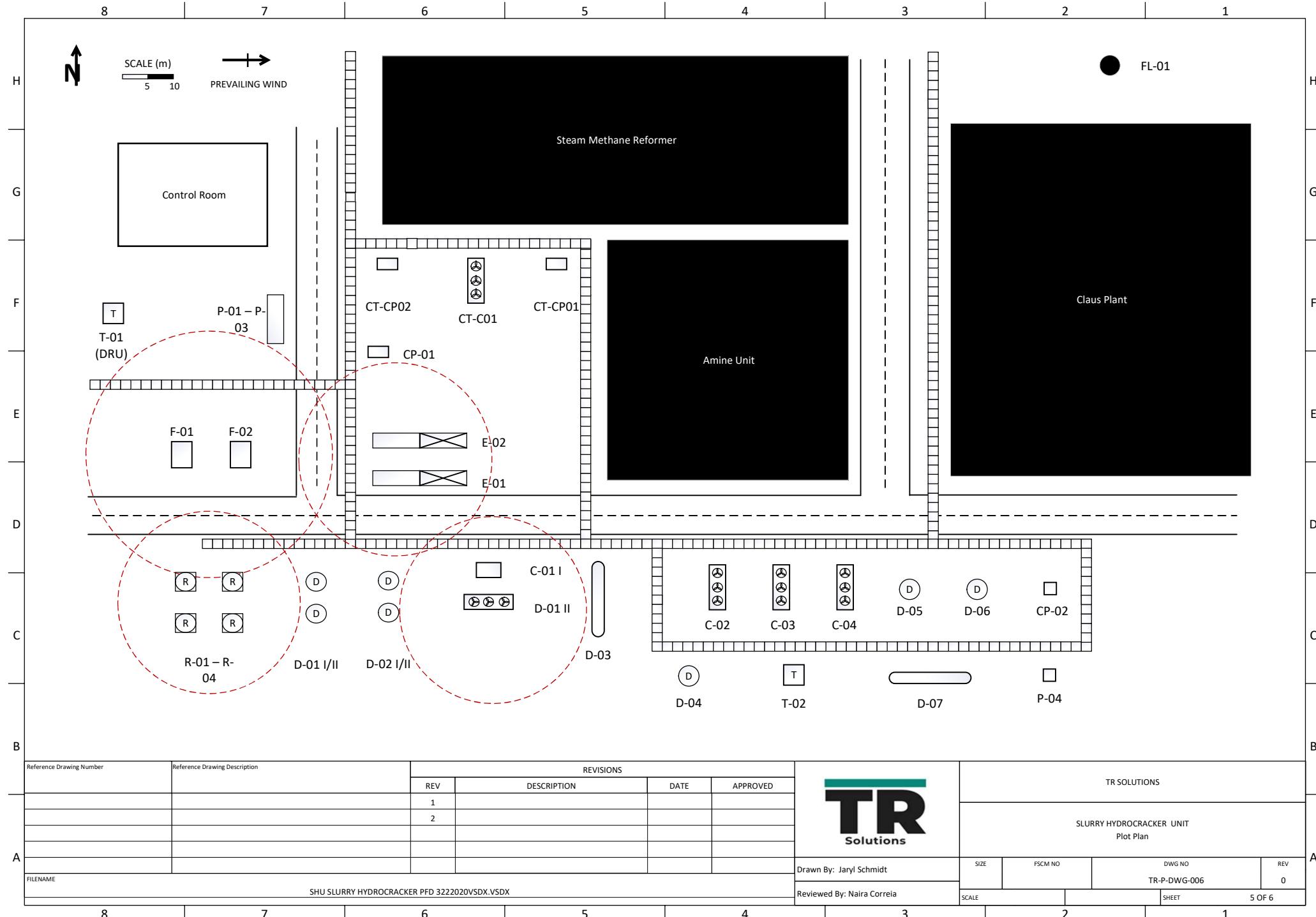
D.4 PLOT PLAN

N

SCALE (m)

5 10

PREVAILING WIND



SLURRY HYDROCRACKER PROJECT

Appendix E - Detailed Equipment Sizing

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E.1 Summary

This appendix contains detailed sizing of major equipment used in the process. Section E2 shows the type, material, specifications, as well as the operating conditions of each of the equipment. Section E3 provides the line sizing table with the pipe diameter, material, and pressure drop of major streams.

E.2 Equipment Sizing

Table E1. Equipment sizing specifications and operating conditions.

Equipment Number	Equipment Description	Equipment Sub-Type	Material	Capacity / Size / Duty Specifications	Temperature (°C)	Pressure (bar)	Mass Flow Rate (tonnes/hr)	Number of Trains	Number of Equipment per Train	Total Number of Equipment
C-01	Waste Heat Boiler	Floating Head	CS/SS	A = 924.5 m ²	380.0	168.0	646.0	2	4	8
C-02	Cooler	Air fin	Stainless Steel	A = 1827 m ²	455.3	79.0	13.0	2	1	2
C-03	Cooler	Air fin	Stainless Steel	A = 84 m ²	319.0	1.0	9.3.0	1	1	1
C-04	Cooler	Air fin	Stainless Steel	A = 1638 m ²	173.0	20.0	630.0	1	1	1
CP-01	Compressor	Centrifugal	Stainless Steel	w _s = 480 kW	70.0	167.0	140.0	1	1	1
CP-02	Compressor	Centrifugal	Stainless Steel	w _s = 9,581 kW	76.0	0.9	33.0	1	1	1
CT-C01	Cooler	Air fin	Carbon Steel	A = 161 m ²	245.3	60.0	40.0	1	1	1
CT-CP01	Compressor	Centrifugal	Stainless Steel	w _s = 26,777 kW	80.0	21.0	40.0	1	1	1
CT-CP02	Compressor	Centrifugal	Stainless Steel	w _s = 27,957 kW	80.0	59.0	40.0	1	1	1
D-01	Flash Drum	Vertical	Stainless Clad	D = 4 m L = 20 m	450.0	169.0	211.0	2	2	4
D-02	Flash Drum	Vertical	Stainless Clad	D = 4 m L = 20 m	380.0	168.0	208.0	2	2	4
D-03	3 Phase Separator	Horizontal	Stainless Clad	D = 3 m L = 7.08 m	70.0	167.0	646.0	1	1	1
D-04	Flash Drum	Vertical	Stainless Clad	D = 1.5 m L = 7.5 m	350.0	78.0	12.7.0	1	1	1
D-05	Flash Drum	Vertical	Stainless Clad	D = 4 m L = 20 m	80.0	20.0	156.0	1	4	4

Equipment Number	Equipment Description	Equipment Sub-Type	Material	Capacity / Size / Duty Specifications	Temperature (°C)	Pressure (bar)	Mass Flow Rate (tonnes/hr)	Number of Trains	Number of Equipment per Train	Total Number of Equipment
D-06	Flash Drum	Vertical	Stainless Clad	D = 4 m L = 20 m	78.2	4.0	305.0	1	2	2
D-07	3 Phase Separator	Horizontal	Stainless Clad	D = 1 m L = 1.1 m	80.0	0.9	93.0	1	1	1
E-01	Heat Exchanger	Floating Head	CS/SS	A = 895 m ²	250	171.0	665.5	2	1	2
E-02	Heat Exchanger	U Tube	CS/SS	A = 30 m ²	72.7	171.0	140.0	2	1	2
EX-01	Gas Expander	Axial	Stainless Steel	w _s = 2640 kW	70.0	167.0	60.0	1	1	1
F-01	Heater	Fired Heater	Carbon Steel	Q = 31627 kW	250.0	171.0	665.5	2	1	2
F-02	Heater	Fired Heater	Carbon Steel	Q = 56431 kW	72.7	171.0	140.0	2	1	2
P-01A	Pump	Centrifugal	Cast Steel	w _s = 769 kW Q = 0.104 m ³ /s	80.0	5.0	333.0	2	1	2
P-01B	Pump	Centrifugal	Cast Steel	w _s = 769 kW Q = 0.104 m ³ /s	80.0	61.0	333.0	2	1	2
P-01C	Pump	Centrifugal	Cast Steel	w _s = 769 kW Q = 0.104 m ³ /s	80.0	116.0	333.0	2	1	2
P-02	Pump	Centrifugal	Cast Steel	w _s = 0.85 kW Q = 0.0016 m ³ /s	80.0	0.9	47.0	1	1	1
R-01	Main Reactor	Vertical	Stainless Steel	D = 4 m L = 30.8 m	466.9	170.0	106.0	2	4	8
R-05, R-06	Conversion Reactor	Vertical	Stainless Steel	D = 4 m L = 19.7 m	119.0	55.0	93.0	1	2	1
T-02	Stripping Column	Vertical	Stainless Steel	D = 1.5 m L = 4.04 m	350.0	78.0	12.6	1	1	1

E.2.1 Two-Phase Separator Sizing Sample Calculations



Two-Phase Separation Sizing

Ex) Flash Drum D-01

According to S13 & S14 of the stream table.

$$T = 450^\circ\text{C}, P = 169 \text{ bar}$$

$$\text{Total flow rates: } Q_L = 16.323 \text{ m}^3/\text{h}, Q_G = 12005.798 \text{ m}^3/\text{h}$$

$Q_G > Q_L \rightarrow$ Gas phase is the continuous phase.
Liquid phase is the dispersed phase.

$$\rho_d = 622.690 \text{ kg/m}^3, \rho_c = 69.388 \text{ kg/m}^3$$

After being splitted into two trains and two vessels per train:

$$Q_{d1} = \frac{16.323 \text{ m}^3/\text{h}}{4} \cdot \frac{\text{h}}{3600\text{s}} = 0.83374 \text{ m}^3/\text{s}$$

$$Q_{c1} = \frac{12005.798 \text{ m}^3/\text{h}}{4} \cdot \frac{\text{h}}{3600\text{s}} = 0.001134 \text{ m}^3/\text{s}$$

Assuming a droplet size of $100 \mu\text{m}$, at $P = 169 \text{ bar}$.

$$k_s = 0.025 \text{ m/s} \quad (\text{Campbell, 2014})$$

$$V_{G,\max} = k_s \sqrt{\frac{\rho_d - \rho_c}{\rho_c}} \quad (\text{Campbell, 2014})$$

$$V_{G,\max} = (0.025 \text{ m/s}) \sqrt{\frac{622.690 - 69.388}{69.388}}$$

$$V_{G,\max} = 0.0706 \text{ m/s}$$

$$D_{\min} = \sqrt{\frac{(4/\pi) Q_c}{F_G V_{G,\max}}} \quad (\text{Campbell, 2014})$$



$F_G = 1$ for vertical vessels. (Campbell, 2014)

$$D_{min} = \sqrt{\frac{(4/\pi)(0.001134 \text{ m}^3/\text{s})}{(1)(0.0706 \text{ m/s})}} = 3.8778 \text{ m}$$

$$\Delta P = P - P_{atm} = 169 \text{ bar} - 1 \text{ bar} = 168 \text{ bar}$$

According to Ulrich & Vasudevan, (2004), Table 4.24:

$$\frac{L}{D} = 5.0 \quad \text{for vessels with an internal pressure} > 35 \text{ bar.}$$

$$L_{min} = (5.0)(D_{min}) = (5.0)(3.8778 \text{ m}) = 19.388 \text{ m}$$

Vessels must have a liquid hold-up time of at least 20 mins. (Rehm et al., 2012)

$$\begin{aligned} V_{hold-up} &= (\dot{Q}_d)(20 \text{ mins})\left(\frac{\text{min}}{60 \text{ s}}\right) \\ &= (0.83374 \text{ m}^3/\text{s})(20 \text{ mins})\left(\frac{\text{min}}{60 \text{ s}}\right) \\ &= 1.360 \text{ m}^3 \end{aligned}$$

$$\begin{aligned} V_{min} &= \frac{\pi}{4} D_{min}^2 L_{min} + V_{hold-up} \\ &= \left(\frac{\pi}{4}\right)(3.8778 \text{ m})^2 (19.388 \text{ m}) + 1.360 \text{ m}^3 \\ &= 230.342 \text{ m}^3 \end{aligned}$$

To satisfy all the limits D_{min} , L_{min} , V_{min} & $\frac{L}{D}$, the actual size of the vessel is determined to be:

$$D = 4 \text{ m. } L = 20 \text{ m.}$$

The rest of the two phase separators are sized in the same manner.

E.2.2 Pump Sizing Sample Calculations



Pump Sizing

ex) Pump P-02 (centrifugal)

According to S49 & S50 of the stream table:

$$T = 80^\circ\text{C}, \quad P_s = 90 \text{ kPa}, \quad P_d = 490 \text{ kPa}$$

$$Q = (5.715 \text{ m}^3/\text{h}) \left(\frac{h}{3600s} \right) = 1.5875 \times 10^{-3} \text{ m}^3/\text{s}$$

$$\Delta P = P_d - P_s = 490 \text{ kPa} - 90 \text{ kPa} = 40 \text{ kPa}$$

Centrifugal pumps typically have an efficiency

$\epsilon_i = 50 - 85\%$. (Ulrich & Vasudevan, 2004)

Assuming $\epsilon_i = 0.75$.

$$W_s = \epsilon_i \frac{Q}{\Delta P} \quad (\text{Ulrich & Vasudevan, 2004})$$

$$= (0.75) \left(\frac{1.5875 \times 10^{-3} \text{ m}^3/\text{s}}{40,000 \text{ Pa}} \right)$$

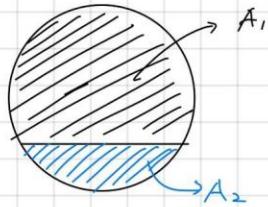
$$= 846.7 \text{ W}$$

The rest of the pumps are sized in the same manner.

E.2.3 Three-Phase Separator Sample Calculations



3 phase separator sizing



- the black shaded area is A_1 , it represents the cross-sectional area occupied by gas phase
- the blue shaded area is A_2 , it represents the cross-sectional area occupied by the two liquid phase.
- for a gas-liquid phase separation, it requires more space than liquid-liquid separation. Thus the sizing will be done based on gas-liquid separation.
- Assume 5 min = 300 s of retention time is needed for both separation and operation requirements.
- If Q represents the volumetric flow rate in m^3/s Then the ratio between A_1 and A_2 will depends on the ratio between Q_{gas} and Q_{liquid} .
- we choose a diameter of $D_V = 3\text{m}$ for this separator, under overall consideration to the scale of this plant.
- Thus $A = A_1 + A_2 = \frac{\pi}{4} D_V^2 = \frac{\pi}{4} 3^2 = 7.068583 \text{ m}^2$

$$A_2 = A \left(\frac{Q_L}{Q_{\text{gas}} + Q_L} \right)$$

- where from simulation:

$$\begin{aligned} \text{For oil phase, } n_{\text{oil}} &= 442967.15 \text{ kg/h} \\ P_{\text{oil}} &= 665.9887 \text{ kPa/m}^2 \end{aligned} \quad \left\{ \begin{array}{l} Q_{\text{oil}} = \frac{n_{\text{oil}}}{P_{\text{oil}}} = 665.13 \text{ m}^3/\text{h} \end{array} \right.$$

$$\begin{aligned} \text{For water phase, } n_{\text{H}_2\text{O}} &= 2608.63 \text{ kg/h} \\ P_{\text{H}_2\text{O}} &= 938.1176 \text{ kPa/m}^2 \end{aligned} \quad \left\{ \begin{array}{l} Q_{\text{H}_2\text{O}} = 2.78 \text{ m}^3/\text{h} \end{array} \right.$$

$$\begin{aligned} \text{For gas phase } n_{\text{G}} &= 200074.56 \text{ kg/h} \\ P_{\text{G}} &= 42.8839 \text{ kPa/m}^2 \end{aligned} \quad \left\{ \begin{array}{l} Q_{\text{G}} = 4665.5 \text{ m}^3/\text{h} \end{array} \right.$$

$$\begin{aligned} \Rightarrow Q_L &= Q_{\text{oil}} + Q_{\text{H}_2\text{O}} = 665.13 \text{ m}^3/\text{h} + 2.78 \text{ m}^3/\text{h} \\ &= 667.91 \text{ m}^3/\text{h} \\ Q_{\text{a}} &= 4665.5 \text{ m}^3/\text{h} \end{aligned}$$



$$- A_2 = A \left(\frac{Q_L}{Q_g + Q_L} \right) = 7.068583 \text{ m}^2 \times \left(\frac{667.90}{667.90 + 1665.50} \right)$$

$$\Rightarrow A_2 = 0.885207 \text{ m}^2$$

- The flow velocity is

$$V_s = \frac{Q_L}{A_2} = \frac{667.90 \text{ m}^3/\text{h}}{0.885207 \text{ m}^2} \times \frac{1\text{h}}{3600\text{s}} = 0.026247 \text{ m/s}$$

- In order to have a retention time for 300 s, that is, the liquid has to flow in the separator for 300 s.

The length of the separator is then

$$L = V_s \times t_R = 0.026247 \text{ m/s} \times 300 \text{ s} = 7.874 \text{ m}$$

- Finally the $\frac{L}{D_r}$ ratio is thereby:

$$\frac{L}{D_r} = \frac{7.874 \text{ m}}{3 \text{ m}} = 2.62$$

which is a normal number.

The sizing for another 3-phase separator is done in the same manner

E.2.4 Control Valve Sizing Sample Calculations



Control Valve sizing

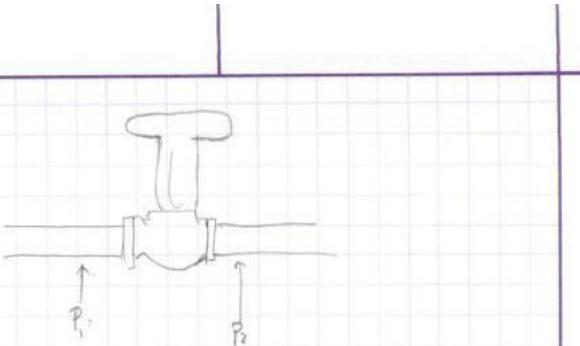
Stream 21

Normal operating conditions

$$T = 70^\circ\text{C}$$

$$Q = 665 \text{ m}^3/\text{h}$$

$$\text{SG} = 0.665$$



Assuming the maximum allowable pressure drop is 2000 kPa across the valve

$$\Delta P = 2000 \text{ kPa}$$

The valve characteristic C_v is described as:

$$C_v = 11.6 Q \sqrt{\frac{\text{SG}}{\Delta P}} \quad (1)$$

$$C_v = 139.1$$

Based on valve selection guidelines, linear opening valves is appropriate with level control with increasing pressure drop as load increases

For proper design, the C_v at normal flow should be at 85% to 95% opening

From catalogue 12 of Fisher Control Valves Page ED-4 (June 2014)

The most appropriate valve has

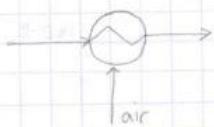
Valve NPS = 6 in ED valve with linear cage
Trim = 7 in part

% travel	C_v
10	463
20	107
30	171
40	228
50	279
60	327
70	367
80	402
90	420
100	433

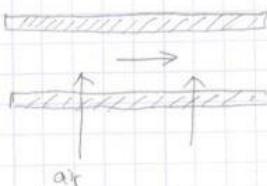
E.2.5 Heat Exchanger Sizing Sample Calculations



1) Detailed Sizing: Heat Exchanger CA-O2



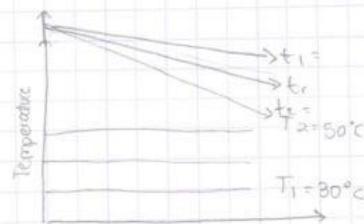
Cross Flow Exchanger



air

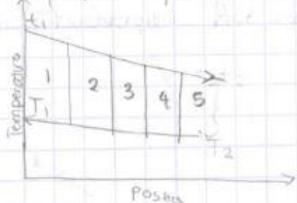
Step 1 Specifications

$$Q_{\text{TOTAL}} = \sum Q_n, n = \text{Zone \#}$$



Position of Hex

The equivalent pseudo-cross-flow temperature-position diagram is shown below.



Where $t_1 = 30^\circ\text{C}$ {assumptions}

$t_2 = 50^\circ\text{C}$

$T_1 = 240^\circ\text{C}$

$T_2 = 70^\circ\text{C}$

The overall mean temperature difference can be calculated using the equation

$$\text{LMTD}_{\text{e}} = \frac{\Delta t - \Delta t_{\text{it1}}}{m \left[\frac{\Delta t}{\Delta t_{\text{it1}}} \right]} = 97.20^\circ\text{C}$$

$$\text{CMTD} = \text{LMTD} \times F$$

According to Hudson Manufacturer's Specifications for MTD correction
(reference ...)

$$R = \frac{T_1 - T_2}{t_2 - t_1} = \frac{240^\circ\text{C} - 70^\circ\text{C}}{50^\circ\text{C} - 20^\circ\text{C}} = 5.66 \quad r = \frac{t_2 - t_1}{T_1 - T_2} = \frac{50 - 30}{240 - 30} = \frac{50 - 30}{240 - 30}$$

for a 1 pass cross flow heat exchanger

$$F \approx 0.95$$

for a 2 pass cross flow

$$F \approx 1$$



$$\therefore \text{CMTD}_t = \text{LMTD} \times F$$

$$= \text{LMTD}$$

$$= 97.20^\circ\text{C}$$

Overall heat transfer coefficient

From Table 4.15b from Ulrich and Vasudevan (2004)

$$U_{\text{air/medium}} = 310 \text{ J/m}^2 \cdot \text{s} \cdot \text{K}$$

hydrocarbons
steam

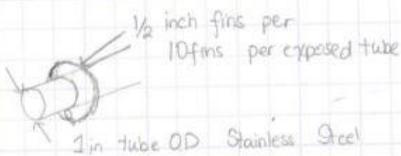
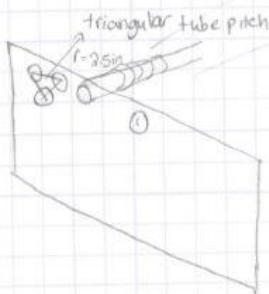
Heat Transfer Area

$$A = \frac{Q}{U_h \text{ LMTD}} = \frac{(110 \times 10^3 \text{ W})}{(310 \text{ J/m}^2 \cdot \text{s} \cdot \text{K})(97.20^\circ\text{C})} = 3450 \cdot 6 \text{ m}^2$$

110W
17.17 h

Finned Tube Dimensions

The selected tube sizing is



To find # of tube rows calculate

$$Z = \left[\frac{T_2 - T_1}{T_i - t_1} \right] = \frac{240 - 70}{240 - 30} = \frac{170}{210} = 0.8091$$

From Table 1 in Hudson (1966)

$$\begin{aligned} \# \text{ of rows} &= 8 \text{ rows} \\ \text{with face velocity} &= 400 \text{ ft/min} \\ &= 0.03 \text{ m/s air velocity} \end{aligned}$$

↳ transversal pitch of 2.5 in = 0.0635m

$$n (\text{number of tubes per row}) = \frac{12}{2.5} \approx 5 \text{ tubes per row}$$

$$a (\text{area per ft of tube}) = \frac{\pi}{12} \cdot 0.03 = 0.2618 \text{ ft}^2/\text{ft} = 0.08 \text{ m}^2$$



The heat transfer rate of fluid is

$$C_f = \frac{2.75 \times 10^7 \text{ W}}{240 - 70} = 161764 \frac{\text{W}}{\text{K}}$$

$$C_{air} = \frac{2.75 \times 10^7 \text{ W}}{50 - 30} = 1375000 \frac{\text{W}}{\text{K}}$$

To size the ACH_E the values of R and k must be calculated to refer to Figure (12) (Hudson's)

$$R = \frac{C_{min}}{C_{max}} = \frac{C_{air}}{C_f} = \frac{1375000}{161764} \frac{\text{W/K}}{\text{W/K}} = 0.118$$

$$NTU = \frac{n \cdot N \cdot a \cdot W \cdot L}{\left[\frac{Q}{T_1 - T_2} \right] \cdot U}$$

$$\text{since } R = \frac{Q}{FV \cdot L \cdot W \cdot 1.08(T_1 - T_2)} \quad (\text{Hudson})$$

$$k = R \cdot NTU = \frac{n \cdot N \cdot a}{1.08 \cdot FV \cdot (1/U)} =$$

$n = 5$ tubes / row

$N = 8$ tubes rows

$a = 0.2(6.8 \text{ ft}^2/\text{ft} \cdot \text{tube})$

$FV = 400 \text{ scfm}$

$U = 545 \text{ BTU} = 310 \text{ W/m}^2\text{K}$

$$k = R \cdot NTU = 1.32$$

$$Z = \frac{T_1 - T_2}{T_1 - t_1} = 0.77$$

Reading from Figure 13 for 2 pass cross flow ACH_E

$$R = 0.5$$

The face Area FA is

$$FA = \frac{R \cdot Q}{(T_1 - T_2)1.08 \cdot FV} = \boxed{187 \text{ ft}^2}$$

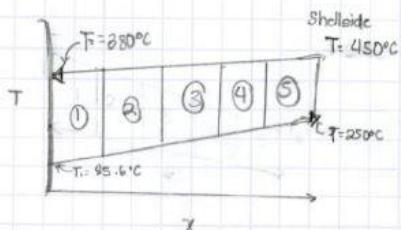
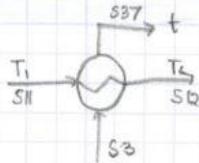


Heat Exchanger Sizing (E-02)

Step 1: Determine $\dot{m} Q$

$$\dot{m} Q = \dot{m} \Delta H = \dot{m} (H_2 - H_1)$$

Stream No.	$H(\text{kJ}/\text{mol})$	Q
1	39415	1463
2	40879	1463
3	42343	1463
4	43807	1463
5	45270	1463



LMTD

$$\text{LMTD} = \frac{\Delta T_i - \Delta T_{i+1}}{\ln \left[\frac{\Delta T_i}{\Delta T_{i+1}} \right]} \quad i.e. \quad \text{LMTD} = \frac{(395 - 122.3) - (380 - 85.6)}{\ln \left[\frac{395 - 122.3}{380 - 85.6} \right]}$$

$$= 260.23^\circ\text{C}$$

Stream	LMTD ($^\circ\text{C}$)
1	282.45
2	260.23
3	240.73
4	223.08
5	207.40

Overall U Value Coefficient

- From Table 9.15a, shell and tube exchange overall coefficient for "asphalt" and condensing vapor stream hydrocarbons with inert gas.

$$U = 100 - 200 \frac{\text{W}}{\text{m}^2 \text{K}} \quad \text{Choose } 150 \frac{\text{W/m}^2}{\text{K}}$$

$$A = \frac{Q_n}{U \text{ LMTD}} \quad A = \frac{1463}{(150)(260.23)} = 574.86 \text{ m}^2$$

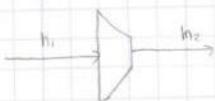
$$\sum A_i = 1790 \text{ m}^2$$

E.2.6 Gas Expander Sizing Sample Calculations



Gas Expander Design

The work outputted by pressure recovery expander can be described by the enthalpy change of inlet and outlet streams



$$W = \dot{m}(h_1 - h_2)$$

$$\dot{m} = 30.01 \text{ kg/h}$$

Using simulation to retrieve specific enthalpy of hydrocarbon stream at $P = 16 \text{ MPa}$ and $P = 3.5 \text{ MPa}$, the isentropic ideal work is

$$W = \frac{\dot{m}}{h} \left(10218 \frac{\text{kJ}}{\text{kmol}} - 7435 \frac{\text{kJ}}{\text{kmol}} \right) \left(\frac{\text{kmol}}{7.48 \text{ kg}} \right) \frac{\text{h}}{3600 \text{ s}}$$

$$= 3.10 \times 10^6 \text{ W}$$

from UOP Figure 4.2 Page 121, the efficiency of gas expander at pressure ratio of 3.02 is approximately 80-85% for power produced

$$\epsilon_i = 85\%$$

\therefore actual power converted

$$W_p = \epsilon_i W_i = 0.85 (3.10 \times 10^6 \text{ W})$$

$$= 2.64 \times 10^6 \text{ W}$$

E.2.7 PSV Sizing Sample Calculations



Pressure Safety Valve Sizing

Based on GPKA Section 5 the Critical Flow through PSV

$$A = \frac{100W\sqrt{T_c(z)}}{C_1 \cdot K \cdot P_i \cdot K_B \cdot \sqrt{MW}}$$

$$C_1 = 387 \sqrt{K \left(\frac{2}{K+1} \right)^{\frac{K+1}{K-1}}} \quad (K=0.975)$$

$$C_1 = 255$$

$$W = 200,074 \text{ kg/hr}$$

$$T_c = 70^\circ\text{C}$$

$$Z = 1.0215$$

$$MW = 7.48 \text{ kg/kmol}$$

$$P_i = 18,300 \text{ kPa} \quad (\text{from MAWP calculation})$$

$$K_B = 1.0$$

$$K = \frac{C_p}{C_v} = 1.112 \quad (\text{from Symmetry})$$

$A = 20.6 \text{ mm}^2$ The designation M with an orifice area of 23.2 mm² is sufficient
 $100 \times 150 \text{ mm}$
 $4" \times 6"$

The PSV is located top of D-03

E.2.8 Reactor Sizing Sample Calculations



Slurry Bubble Reactor Design and Sizing

In order to completely design the Slurry reactor some kinetics would need to be provided, therefore, Mr. Nolte provide the CANMET reactor size specification:

$$\text{Space velocity LHSV} = 0.21/\text{hr}$$

$$\text{Vapor voidage} = 40\%$$

From simulation:

$$\dot{V}_{in} = 15598 \text{ m}^3/\text{day}$$

$$\dot{m}_{in} = 665516 \text{ kg/hr}$$

$$\text{MW} = 498.91 \text{ kg/kmol}$$

Assumptions:

- Reactor conversion 95% : $X_A = 0.95$
- Continuous stirred-tank reactor (CSTR)
- well mixed
- steady-state reaction throughout the reactor
- First-order reaction

$$\text{LHSV} = \frac{\text{volumetric feed liquid flow rate}}{\text{reaction volume}} = \frac{\dot{V}}{V}$$

$$V = 15598 \frac{\text{m}^3}{\text{day}} \times \frac{1 \text{ day}}{24 \text{ hr}} \times \frac{1 \text{ hr}}{0.21} = 2094.8 \text{ m}^3$$

Overall Mole Balance:

$$\text{In - Out + Generation - Consumption} = \text{Accumulation} \quad \nearrow t^0$$

$$F_{A0} - F_A + \int r_A dV = 0$$

$$F_{A0} - F_A + r_A V = 0$$

$$r_A = \frac{F_A - F_{A0}}{V}$$



Mole in:

$$\dot{F}_{A0} = \frac{\dot{m}}{MW} = 665516 \frac{\text{kg}}{\text{hr}} \times \frac{\text{kmol}}{498.91 \text{ kg}} \times \frac{\text{mol}}{1000 \text{ kmol}} = 1.33 \frac{\text{mol}}{\text{hr}}$$

Mole out:

$$\dot{F}_A = \dot{F}_{A0} (1 - X_A) = 1.33 \frac{\text{mol}}{\text{hr}} \times (1 - 0.95) = 0.067 \frac{\text{mol}}{\text{hr}}$$

Rate law for CSTR 1st order reaction:

$$-\dot{r}_A = K C_A$$

Initial concentration:

$$C_{A0} = \frac{\dot{F}_{A0}}{V} = 1.33 \frac{\text{mol}}{\text{hr}} \times \frac{\text{day}}{15598 \text{ m}^3} \times \frac{24 \text{ hr}}{1 \text{ day}} = 2.05 \times 10^{-3} \frac{\text{mol}}{\text{m}^3}$$

Final concentration:

$$C_A = C_{A0} (1 - X_A) = 2.05 \times 10^{-3} \frac{\text{mol}}{\text{m}^3} \times (1 - 0.95) = 1.03 \times 10^{-4} \frac{\text{mol}}{\text{m}^3}$$

$$\dot{r}_A = \frac{\dot{F}_A - \dot{F}_{A0}}{V} = \frac{0.067 - 1.33}{3094.8} = -4.08 \times 10^{-4} \frac{\text{mol}}{\text{m}^3 \cdot \text{hr}}$$

$$-\dot{r}_A = K C_A \Rightarrow \text{Rate constant } K = -\frac{\dot{r}_A}{C_A}$$

$$K = -\frac{(-4.08 \times 10^{-4})}{1.03 \times 10^{-4}} = 3.96 \text{ hr}^{-1}$$

Half-life for 1st order

$$t_{1/2} = \frac{0.693}{K} = \frac{0.693}{3.96} = 0.17 \text{ hr}$$

Residence time

$$\tau = \frac{C_{A0} V}{\dot{F}_{A0}} = \frac{2.05 \times 10^{-3} \times 3094.8}{1.33}$$

$$\tau = \underline{\underline{4.77 \text{ hr}}}$$



Due to the large amount of volume the residence time is very large. Therefore, the technical decision was to have 2 trains with 4 identical reactors in series.

Giving in total 8 Slurry Bubble Reactor operating at 250 bar and 467 °C with a volume of

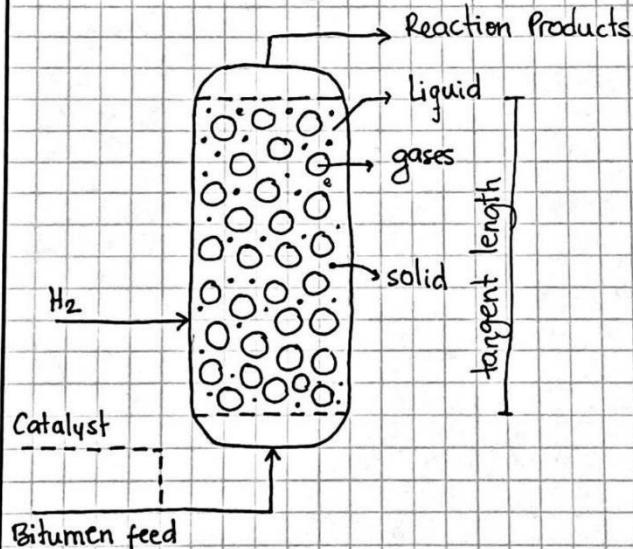
$$V = \frac{3094.8 \text{ m}^3}{8} = 386.9 \text{ m}^3 \text{ per each reactor.}$$

New Residence time for each reactor

$$\tau = \frac{2.05 \times 10^{-3} \times 386.9}{1.33}$$

$$\tau = 0.6 \text{ hr}$$

Reactor Sizing R-01



Darby equation:

$$u_t = \frac{\mu_f}{\rho_f D_p} \left[\left(14.4 + 1.8 \sqrt{D_p^3 (\rho_p - \rho_f) \rho_f g / \mu_f^2} \right)^{0.5} - 3.8 \right]^2$$



$$\text{Particle size : } D_p = 100 \mu\text{m} = 1 \times 10^{-4} \text{ m}$$

$$\rho_{\text{liquid}} = 527.02 \text{ kg/m}^3$$

$$\rho_{\text{gas}} = 99.86 \text{ kg/m}^3$$

$$\mu_{\text{liquid}} = 3.78 \times 10^{-6} \text{ Pa.s}$$

$$g = 9.81 \text{ m/s}^2$$

$$u_t = \frac{3.78 \times 10^{-6}}{527.02 \times 1 \times 10^{-4}} \left[14.4 + 1.8 \sqrt{\frac{(1 \times 10^{-4})^3 (527.02 - 99.86) \cdot 527.02 \times 9.81}{(3.78 \times 10^{-6})^2}} \right]^{0.5} - 3.8 \text{ m/s}$$

$$u_t = 0.0382 \text{ m/s}$$

For vertical drum

$$D = \left(\frac{5 \dot{q}}{\pi u_t} \right)^{0.5}$$

$$D = \left(5 \times \frac{15598 \text{ m}^3}{\text{day}} \times \frac{1 \text{ day}}{86400 \text{ s}} \times \frac{1}{0.0382 \text{ m}} \times \frac{1}{\pi} \right)^{0.5}$$

$\underline{D = 2.74 \text{ m}}$ \Rightarrow minimum diameter per each reactor

$$V = \pi r^2 L = \frac{\pi}{4} D^2 L \quad \Rightarrow \quad L = \frac{4V}{\pi D^2}$$

$$L = \frac{4 \times 386.9 \text{ m}^3}{\pi \times (2.74 \text{ m})^2} = 65.6 \text{ m}$$

Check L/D ratio : (Gerundia, 1981)

Based on Pressure of the drum 250 bar-a $L/D \geq 5$

$$\frac{L}{D} = \frac{65.6 \text{ m}}{2.74 \text{ m}} = 23.9$$



The L/D ratio is too high, therefore increase the diameter in order to obtain a better ratio

$$D = 4 \text{ m}$$

$$L = \frac{4 \times 386.9 \text{ m}^3}{\pi \times (4)^2 \text{ m}^2} = 30.8 \text{ m}$$

$$\Rightarrow \frac{L}{D} = \frac{30.8 \text{ m}}{4 \text{ m}} = 7.7 \Rightarrow \text{more acceptable ratio!}$$

$$\Rightarrow \frac{L}{D} = 7.7, D = 4 \text{ m} \cancel{/}$$

~~.....~~

Pressure vessel wall thickness

$$t = \frac{Pr}{SE - 0.6P} + C_s$$

$$r = \frac{4 \text{ m}}{2} = 2 \text{ m}$$

S = maximum allowable tensile stress

U&V Fig 4.1 : \rightarrow Material required : carbon steel

$$\rightarrow T = 467^\circ\text{C}$$

$$S = 840 \text{ bar}$$

E = welded joint efficiency $E = 0.9$

C_s = corrosion allowance $C_s = 3 \text{ mm}$ for carbon steel

$$t = \frac{250 \times 2}{(840 \times 0.9) - (0.6 \times 250)} + 0.003 = 0.828 \text{ m}$$

$$t = 828 \text{ mm} \cancel{/}$$

E.2.9 Stripper Column Sizing Sample Calculations



Stripper Column Sizing

In order to design the tower, the diameter and height must be determined.

From Symmetry

$$N_{\text{theoretical}} = 2 \text{ trays}$$

L = Liquid molar flowrate (mol/h)

V = Gas molar flowrate (mol/h)

M_g = molar mass of gas (g/mol)

ρ_L = liquid density (kg/m³)

ρ_V = gas density (kg/m³)

Diameter of the tower

$$D = \sqrt{\frac{4V_{\text{max}} M_g}{\pi \rho_g \cdot U_{\text{sg}}}} \quad (1)$$

where

$$U_{\text{sg}} = k_{\text{SB}} \sqrt{\frac{\rho_L - \rho_V}{\rho_V}} \quad (2)$$

Based on Ulrich and Vasudevan (2004), a value of 0.06 (lower end) applies to gas-liquid systems

$$U_{\text{sg}} = 0.06 \text{ m/s} \sqrt{\frac{919 \text{ kg/m}^3 - 0.7 \text{ kg/m}^3}{0.7 \text{ kg/m}^3}} = 2.17 \text{ m/s}$$

$$D = \sqrt{\frac{4(270.34 \text{ mol/m}) (34.26 \text{ kg/mol}) (1/\text{m}^3) (3600 \text{ s})}{\pi (0.699 \text{ kg/m}^3) (2.17 \text{ m/s})}} = 1.5 \text{ m}$$

Tray efficiency can be estimated at 80% from (Nag, Ashish) (2016) as conservative option. This was obtained by assuming liquid is similar to HClO product.

Based on CH E 464 lecture, default size spacing of 24 in can be used.

$$H = \frac{N_{\text{theoretical}} \times H_t}{E_s}$$



Thus,

$$H = \frac{(2)}{0.3} \cdot (24\text{in}) \left(\frac{0.0254\text{m}}{1\text{in}} \right) = 1.04\text{m}$$

The liquid holdup is also part of total tower height

$$H_{LH} = \frac{V_{al}}{A} \xrightarrow{\text{→ Volumetric flowrate at last tray}} \text{liquid residence time :}$$

$$H_{LH} = \frac{(7.4\text{ m}^3/\text{h})(0\text{ min}) (1/60)}{(\pi)(1.5\text{ m}^2)} = 0.204\text{ m}$$

Dimensions

$$D = 1.5\text{m}$$

$$H_t = 1.04\text{m}$$

$$\text{Tray spacing} = 24\text{in}$$

$$N_{actual} = \frac{2}{0.3} = 7\text{ trays}$$

E.3 Line Sizing

Table E2. Detailed line designation.

Stream Number	Nominal Pipe Diameter (in)	Pipe Schedule	Velocity (m/s)	Pressure Drop (kPa/100m)	Material	Material Grade	Type of Flow	Flow Regime
S3	8	20	3.1	35	ASTM A106 (CS)	B	Liquid	Turbulent
S4	8	100	3.7	55	ASTM A106 (CS)	B	Liquid	Turbulent
S5	8	100	4.6	68	ASTM A106 (CS)	B	Liquid	Turbulent
S6	8	100	5.8	83	ASTM A106 (CS)	B	Liquid	Turbulent
S7	4	160	0.0	n/a	ASTM A106 (CS)	B	Solid	Turbulent
S8	8	160	5.9	84	ASTM A106 (CS)	B	Slurry	Turbulent
S9	14	120	24.2	19	ASTM A376 (SS)	TP304	Vapour	Turbulent
S10	4	140	0.0	n/a	ASTM A376 (SS)	TP304	Slurry	Turbulent
S11	18	140	15.2	26	ASTM A376 (SS)	TP304	Vapour	Turbulent
S12	18	120	14.2	29	ASTM A376 (SS)	TP304	2-Phase	Dispersed
S13	1.5	80	2.0	51	ASTM A376 (SS)	TP304	Liquid	Turbulent
S14	18	140	14.9	25	ASTM A376 (SS)	TP304	Vapour	Turbulent
S15	18	120	12.8	32	ASTM A376 (SS)	TP304	2-Phase	Dispersed
S16	8	120	2.0	5	ASTM A376 (SS)	TP304	Liquid	Turbulent
S17	16	120	15.6	27	ASTM A376 (SS)	TP304	Vapour	Turbulent
S18	16	120	12.1	34	ASTM A376 (SS)	TP304	2-Phase	Dispersed
S19	18	100	11.9	39	ASTM A376 (SS)	TP304	3-Phase	Dispersed
S20	1	40	1.4	71	ASTM A376 (SS)	TP304	Liquid	Turbulent
S21	8	100	6.6	84	ASTM A376 (SS)	TP304	Liquid	Turbulent
S22	12	100	20.9	44	ASTM A376 (SS)	TP304	Vapour	Turbulent
S23	12	100	14.6	21	ASTM A376 (SS)	TP304	Vapour	Turbulent
S24	10	100	10.2	13	ASTM A376 (SS)	TP304	Vapour	Turbulent
S25	10	120	20.3	28	ASTM A376 (SS)	TP304	Vapour	Turbulent

E.3.1 Line Sizing Sample Calculations

LINE SIZING

General Assumptions / Notes:

- * at points where streams merge/split between parallel pieces of equipment (e.g. R-A01 - R-A04), the largest combined flow rate is used for line sizing
- * $\frac{1}{16}$ in corrosion allowance was applied to carbon steel lines; $\frac{1}{32}$ in corrosion allowance applied to stainless steel lines
- * 100m characteristic length used for all sizing
- * line sizing was not performed for S7 (solid catalyst input) or S10 (reactor bottoms slurry line) as they would require more detailed analysis

Pipe Wall Thickness (Pipe Schedule)

Pipe wall thickness for allowable working pressure determined by ANSI B31.3, "Code for Pressure Piping, Petroleum Refinery Piping"

$$t = \frac{P_{do}}{2(Se + Ry)} \quad (\text{GPSA, Fig 17-23})$$

* data for allowable material stress, S_e , obtained from GPSA, Fig 17-25

Example: S8 = set nominal pipe size to 8.0in

$$t = \frac{(17,000 - 101.325)(219.0\text{mm})}{2[117\text{kPa} \cdot 1 + (17,000 - 101.325) \cdot 0.4]} = 14.8 \text{ mm}$$

$$t_m = 14.8 \text{ mm} + 1.59 \text{ mm} = \boxed{16.3 \text{ mm} \equiv \text{minimum pipe wall thickness}}$$

design thickness corrosion allowance

@ 8.0in pipe, closest schedule pipe is schedule 100, $t = 18.3\text{mm}$

Liquid Flows

Pressure loss due to friction is calculated using the Darcy-Weisbach Equation

$$\boxed{\Delta P_f = \frac{0.5 f_m L V^2}{d}} \quad (\text{GPSA, Eq 17-2})$$

* Moody friction factor, f_m , obtained from GPSA, Fig 17-2

Example: 56

$$\Delta P_f = \frac{0.5 (606.27 \text{ kg/m}^3) (0.0146) (100\text{m}) (5.835 \text{ m/s})}{1982.4 \text{ mm}} = \boxed{83 \text{ kPa}/100\text{m} = \Delta P_f}$$

Vapour Flows

Pressure loss due to friction (for vapour flows) is calculated using a simplified Darcy-Weisbach formula.

$$\boxed{\Delta P_{100} = \frac{W^2}{e} \left[\frac{62,350 (10^4) F}{d^5} \right]} \quad (\text{GPSA, Eq 17-30})$$

which can be simplified to:

$$\boxed{\Delta P_{100} = \frac{C_1 C_2}{e}} \quad (\text{GPSA, Eq. 17-31})$$

where: $C_1 = W^2 (10^4)$, obtained from GPSA, Fig 17-8

$$C_2 = \frac{62,350 \cdot 10^4 \cdot F}{d^5}, \quad \text{obtained from GPSA, Fig 17-9}$$

Example: 59

- W (mass flow rate) = $90,114 \text{ kg/h} \Rightarrow$ Fig 17-8 $\Rightarrow C_1 = 8.4$

- Pipe used: 14.0 in, schedule 120 \Rightarrow Fig 17-9 $\Rightarrow C_2 = 33.40$

$$\Delta P_{100} = \frac{(8.4)(33.40)}{14.58 \text{ kg/m}^3} = \boxed{19 \text{ kPa}/100\text{m} = \Delta P_{100}}$$

Two-Phase Flow

The Dukler equation was used for frictional pressure drop calculations

$$\Delta P_{fr} = \frac{f_n F_{tpr} \rho_w V_m^2 L_m}{2d} \quad (\text{GPSA, Eq 17-30})$$

where $F_n = 0.0056 + 0.5/(Re_y)^{-0.32}$ (GPSA, Eq 17-44)

$$Re_y = \frac{(0.001) \rho_w V_m d}{\mu_w} \quad (\text{GPSA, Eq 17-45})$$

$$F_{tpr} = \frac{3.37}{1.281 - 0.478y + 0.444y^2 - 0.094y^3 + 0.00843y^4 + 1} \quad (\text{GPSA, Eq 17-48})$$

$$y = -\ln(\lambda)$$

Example: S15

$$y = -\ln(0.0343) = \underline{\underline{3.37}} = y$$

$$F_{tpr} = \frac{3.37}{1.281 - 0.478(3.37) + 0.444(3.37)^2 - 0.094(3.37)^3 + 0.00843(3.37)^4 + 1} \quad \underline{\underline{f_{tpr} = 2.53}}$$

$$Re_y = \frac{(0.001)(76.96 \text{ kg/m}^3)(387.2 \text{ mm})}{17510^{-5} \text{ Pa}\cdot\text{s}} = \underline{\underline{2.17 \cdot 10^7}} = Re_y$$

$$F_n = 0.0056 + 0.5(2.17 \cdot 10^7)^{-0.32} = 7.98 \cdot 10^{-3} = F_n$$

$$\Delta P_{fr} = \frac{(7.98 \cdot 10^{-3})(2.53)(76.96 \text{ kg/m}^3)(12.77 \text{ m/s})^2 (100 \text{ m})}{2(387.2 \text{ mm})} = \boxed{\underline{\underline{32 \text{ kPa}/100 \text{ m} = \Delta P_{fr}}}}$$

E.4 References

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SLURRY HYDROCRACKER PROJECT

Appendix F - Environmental Assessment

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DATE: April 9th, 2020

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F.1 SUMMARY

This appendix shows the environmental analysis of the emissions during the conversion of the slurry bitumen and sample calculation.

F.2 ENVIRONMENTAL ANALYSIS

Due to the combustion of the PSA off-gas, the Claus plant operation, and natural gas burning, this project emits a total amount of 4.77 megatons of greenhouse gases (GHG) annually. Indirect GHG associated with electricity consumption due to electric motors of the pumps, compressors, and expander is 0.52 megatons per year. In summation, this project results in 5.29 megatons of annual carbon footprint. The detailed break-down is illustrated in Table F1.

Table F 1: Annual Emission from different sources.

Annual Emissions				
Direct	PSA Vent Gas Combustion	CO ₂	2.43	megatons
		H ₂ O	0.75	megatons
	Claus Plant	CO ₂	0.0015	megatons
		H ₂ O	0.12	megatons
	Natural Gas Fuel	CO ₂	0.81	megatons
		H ₂ O	0.66	megatons
		NO ₂	<7.8E-04	ppb in ambient air
		SO ₂	<8.5E-06	ppb in ambient air
		Particulates	<1.2E-13	µg/m ³ ambient air
Indirect	Electricity Consumption	CO ₂	0.52	megatons

F.2.1 Emission Sample Calculations



Emission Sample Calculations

ex) GHG emissions due to H₂ combustion

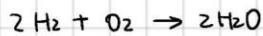
According to VMGI Symmetry, in PSA vent gas (S35)

Total molar flow rate = 10551.84 kmol/hr

Mol % H₂ = 0.3333

$$\begin{aligned}\therefore \text{Molar flow rate of H}_2 &= (0.3333)(10551.84 \text{ kmol/hr}) \\ &= (3516.93 \text{ kmol/hr}) \left(\frac{24 \text{ hr}}{\text{day}} \right) \left(\frac{365 \text{ days}}{1 \text{ yr}} \right) \\ &= 3.081 \times 10^7 \text{ kmol/yr}\end{aligned}$$

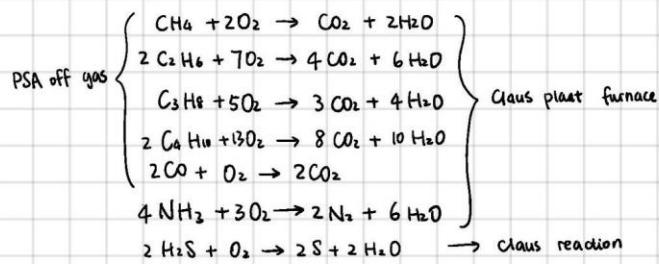
Combustion Reaction kinetics of H₂:



Molar flow rate of H₂O = $3.081 \times 10^7 \text{ kmol/yr}$

$$\begin{aligned}\text{Mass flow rate of H}_2\text{O} &= (3.081 \times 10^7 \text{ kmol/yr}) \left(\frac{18.015 \text{ kg}}{\text{kmol}} \right) \\ &= (5.55 \times 10^8 \text{ kg/yr}) \left(\frac{\text{ton}}{1000 \text{ kg}} \right) \left(\frac{\text{megaton}}{10^6 \text{ ton}} \right) \\ &= 0.555 \text{ megaton/yr}\end{aligned}$$

The rest of GHG emissions are calculated in the same manner with the following reaction kinetics:



ex) Indirect GHG emissions due to electricity consumption from the Utility consumptions (Appendix I)

Total electricity consumption = $5.29 \times 10^8 \text{ kWh/yr}$

According to National Energy Board (2019), in Alberta, every kWh of electricity generated is equivalent to 790 g of CO₂ emitted.

$$\begin{aligned}\text{Indirect CO}_2 \text{ emission} &= (790 \text{ g/kWh}) (5.29 \times 10^8 \text{ kWh/yr}) \left(\frac{\text{kg}}{1000 \text{ g}} \right) \left(\frac{\text{ton}}{1000 \text{ kg}} \right) \left(\frac{\text{megaton}}{10^6 \text{ ton}} \right) \\ &= 0.42 \text{ megaton CO}_2 / \text{yr}\end{aligned}$$

F.3 REFERENCES

- [1] National Energy Board. (2019, August 28). Canada's Renewable Power Landscape 2017 – Energy Market Analysis. Retrieved from <https://www.cer-rec.gc.ca/nrg/sttstc/lctrct/rprt/2017cntrnwblpwr/ghgmssn-eng.htm>
- [2] “Canadian Ambient Air Quality Standards.” *NaturalGas.org*, naturalgas.org/environment/naturalgas/.

SLURRY HYDROCRACKER PROJECT

Appendix G – Safety & Risk Assessment

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G.1 SUMMARY

This appendix shows the process hazard analyses (PHA). A What-if analysis, Dow's fire and explosion indexes (FEI), and Dow's chemical exposure indexes (CEI), and Material Safety Data Sheets along with the methodologies and assumptions used are discussed.

G.2 WHAT-IF ANALYSIS

The PHA What if analysis was performed using the process PFD and risk matrix. The risk matrix will be limited to the consequence categories of damage to the asset, health, and safety/environmental risk evaluations, and the probability of an event occurring. After the possible causes are evaluated and the possible consequences are ranked based on the level of severity and likelihood then engineering safeguards action are determined to bring the risk to an acceptable level according to the risk matrix.

Table G1. Risk Matrix.

Table G2: What If Analysis

ITEM	GUIDE WORD	SITUATION		BEFORE SAFEGUARDS				AFTER SAFEGUARDS				AFTER RECOMMENDATIONS		PO ST RE CO M ME ND ATI ON LIK ELI HO OD	RIS K RA NK AF TE R RE CO M ME ND ATI ON ME ND ATI ON S
		CAUSE (What if/ What happened?)	CONSEQUENCE (What is the result?)	CONSEQUENCE CATEGORY	CONSEQUENCE (Severity)	LI K E L H O O D	RI S K R A N K	MITIGATING MEASURES (Safeguards in Place and Safeguard Category)	M IT I G A T E D LI K E LI H O O D	R E SI D U A L RI S K R A N K	RECOMMENDATIONS (Additional safeguards or actions required to be implemented to reduce Residual Risk Rank to acceptable level)				
1	High Pressure	Pressure build-up in bitumen storage vessel due to downstream pipe blockage or valve failure.	Vessel can burst, leading to potential fire and explosion hazard.	Assets (MPPD)	6	3	8	A PSV flare brings the conditions to atmospheric pressure in case of a pressure build-up.	2	7	1. Monitor the pressure in the vessel using a pressure indicator. 2. Execute an emergency shutdown of the unit if pressure continues to increase after PSV is opened. 3. Do not resume operation until any downstream obstacles have been removed or a bypass has been used.	1	6		
2	Contamination	Release of hot bitumen.	Can cause thermal burns, formation or release of hydrocarbon fumes	People workers and public	5	2	6	Pressure drop alarm and emergency shutdown of the system can avoid feed contamination	1	5	1. Perform shutdowns and turnarounds on the reactors every three to five years. 2. Implement first aid response. 3. Workers must wear PPE all the time during operation.	1	5		

3	Loss of Containment	Sour Gas and hydrogen sulfide release.	Highly fatal if inhaled, can cause severe injuries	People workers and public	6	2	7	1. If the release happens in one of the trains, a train emergency shutdown will be performed and the rest of the trains can operate normally. 2. If the release happens in the streams without possible bypass, a unit emergency shutdown will be performed.	1	6	1. Implement first aid response in case of emergency. 2. Workers must wear PPE all the time.	1	6
4	High Flow	Higher mass flow rate through the reactor, causing downstream flash drums (D-04, D-05) to operate at over capacity.	1. Separator burst. 2. High pressure and temperature material is released to atmosphere. 3. Fire and explosion hazard.	People workers and public	6	2	7	A train emergency shutdown will be performed and the rest of the trains can operate normally.	1	6	1. Implement first aid response in case of emergency. 2. Workers must wear PPE all the time.	1	6
5	Low/No Flow	Bitumen is fed to the reactors at lower mass flow rates.	Loss of production.	Assets (MPPD)	1	2	2	None.	2	2	Monitor liquid hold-up levels in process vessels to ensure they are in normal range.	1	1
6	Utility Failure	Failure on the bitumen feed furnace fuel gas storage.	An unburned mixture of gas and air can enter the gas fire which can enter the hot fire box and ignite.	Assets (MPPD)	6	2	7	A train emergency shutdown will be performed and the rest of the trains can operate normally.	1	6	Check if fuel gas lines are liquid free. Ensure that the burner and pilot gas valves are fully closed before pulling blinds.	1	6
7	High Flow	Leakage on bitumen feed pump P1.	1. Pump damage. 2. Major release of process chemicals (carbon aromatics and paraffins) 3. Potential fire or explosion hazard.	People workers and public	6	2	7	1. Pressure indicators on pipelines and vessels will alarm when pressure drops. 2. A train emergency shutdown will be performed and the rest of the trains can operate normally.	1	6	1. Perform routine pump inspections and maintenance. 2. Workers must wear PPE all the time. 3. Implement first aid response in case of injury.	1	6
8	Startup / Shutdown Hazards	Pump P1 discharge valve is closed.	Can cause pump failures due to overpressure and mechanical stress.	Assets (MPPD)	4	2	5	1-A pump discharge check valve will prevent backflow. 2-Ensure proper training is provided to the employee operating the equipment.	1	4			

9	Startup / Shutdown Hazards	Pump P1 suction valve is closed	Can cause failure including overpressure and mechanical stress in the pump inlet and possibly causing the pipe to burst	Assets (MPPD)	2	2	3	1- Unit emergency shutdown 2-Install a pressure indicator 3-Ensure proper training is provided to the employee operating the equipment 4-consider installation of isolation valves	1	2	
10	Low / No Flow	Pump P1 operates at low flow rate	Loss of production	Assets (MPPD)	3	2	4	1-Install a low suction pressure alarm 2-Vibration monitor	1	3	
11	High Pressure	Piping downstream to the reactor is blocked and the reactor continues to send high pressure fluid.	1. Piping burst. 2. High pressure and temperature material released to atmosphere, potentially causing explosion.	People workers and public	6	2	7	1. Pressure safety valve monitors and releases pressure inside the reactor to prevent pressure build-up. 2. High pressure alarm on reactors and pipings. 3. Train is shutdown until maintenance is completed. The alternative trains can continue to operate.	1	6	1. Perform routine process equipment inspections and maintenance. 2. Workers must wear PPE all the time. 3. Implement first aid response in case of injury.
12	High Pressure	Upstream equipment delivers higher pressure fluid, causing the distillation columns to operate at over pressure.	1. Reboiler or piping leaks. 2. High pressure and temperature material released to atmosphere, potentially causing explosion.	People workers and public	6	2	7	1. Pressure safety valve monitors and releases pressure inside the reactor to prevent pressure build-up. 2. High pressure alarm on vessels and pipings. 3. Train is shutdown until maintenance is completed. The alternative trains can continue to operate.	1	6	1. Perform routine process equipment inspections and maintenance. 2. Workers must wear PPE all the time. 3. Implement first aid response in case of injury.
13	Reverse or Misdirected flow	Separator (D-08) operates at a lower pressure than downstream units.	Fluid flow backwards from mixer (M-07) to the separator, causing contaminations.	Assets (MPPD)	4	2	5	A pressure check valve at the separator discharge will prevent backflow.	1	4	Monitor the pressure indicators in each process vessel to ensure the pressures are maintained in normal ranges.
14	High Pressure	Excessive pressure on the 3-phase Separator (D-03)	1. Potential leak and explosion hazard. 2. Leakage of dangerous gas exposure (H2S) to the surrounding environment	People workers and public	6	3	8	1. Pressure valve controls 2. Pressure gauges 3. Thermometers 4. Emergency shutdown of the unit.	2	7	1. Wear proper PPE to prevent inhalation of toxic gases 2. Emergency response team on site 3. To avoid the explosion to occur in a domino effect emergency response plan must be evaluated in order to anticipate explosion

1 5	High Level	Increase of flow into separators, causing liquid hold-up height to rise.	Insufficient vapor-liquid disengagement space. Liquid entrained in gas enters compressor, causing damage to the equipment.	Assets (MPPD)	4	2	5	The high level of liquid triggers the level control system of the separators, increasing liquid flow rate and bringing liquid hold-up height back in normal range.	1	4	Monitor the level indicator of the vessels and ensure the liquid level in maintained within normal range.	1	4
1 6	Low level	Decrease of flow into separators, causing liquid hold-up height to drop.	Loss of liquid hold-up at the bottom of the vessel. Unseparated fluid bypasses at the bottom.	Assets (MPPD)	3	2	4	The low level of liquid triggers the level control system of the separators, reducing liquid flow rate and bringing liquid hold-up height back in normal range.	1	3	Monitor the level indicator of the vessels and ensure the liquid level in maintained within normal range.	1	3
1 7	Utility Failure	Sudden power cut to the pump drivers.	1. Reverse flow though the pumps. 2. Internal pump components may be damaged or dislodged.	Assets (MPPD)	4	3	6	A discharge check valve will allow forward flow.	1	4	Perform inspections and maintenance on the pumps after power failures to ensure the internal components have not been dislodged or damaged.	1	4
1 8	Maintenance Hazards	Corrosion in the slurry bubble reactor due to catalyst poisoning and deactivation	Reactor and piping leaking. High pressure and temperature material released to atmosphere, potentially causing explosion.	People workers and public	6	2	7	Perform mantainance around the reactor regularly	1	6	1. Perform mantainance around the reactor regularly. 2. Workers must wear PPE all the time. 3. Implement first aid response in case of injury.	1	6
1 9	High Temperature	Bitumen is fed at a higher temperature than being specified due to operation error.	Reactor does not operate at designed condition. Products may not meet specifications.	Assets (MPPD)	4	1	4	Install temperature indicators on the reactors to monitor temperature change.	1	4	Provide proper training and accessible operating procedure for the workers.	1	4
2 0	Low / No Flow	Pressure safety valve fails to release pressure inside separators.	Pressure build-up in the vessels, causing overpressure operation. Vessel or pipes may eventually burst and release flammable material.	Assets (MPPD)	6	1	6	Emergency shutdown of the train. The alternative trains can continue to operate normally.	1	6	Monitor the pressure indicators of the vessels.	1	6

2 1	Contamination	Leakage in heat exchanger EX-01. Product gas is contaminated with unreacted feed fluid.	Final product fail to meet specifications.	Assets (MPPD)	4	2	5	1. Stainless clad material is used for the heat exchangers to avoid corrosion. 2. A composition indicator can be installed to monitor product compositions.	1	4	Perform routine inspections and maintenance on the heat exchangers.	1	4
2 2	Loss of Containment	Storage tank overfill.	1. Loss of material. 2. Spill of flammable material to the working environment.	Assets (MPPD)	3	3	5	None.	3	5	1. Provide proper training and accessible operating procedure for the workers. 2. Perform management of changes. Expand the storage unit when needed.	1	3
2 3	Contamination	Water comes in contact with bitumen	1. Corrosion to the process equipment. 2. Moisture buildup in the tank insulation 3. Potential spills and falls	Assets (MPPD)	3	1	3	Stainless clad material is used for most of the process equipment to prevent corrosion.	1	3	Monitor the composition of the process fluid. In case of excess moisture, a potential dehydration strategy is the installation of a glycol dehydration unit.	1	3
2 4	High Level	Blockage in liquid discharge stream of the reactor or increase in the reactor feed flow.	1. Liquid accumulation in the column. 2. Reduced gas hold up volume in the reactor. 3. Loss of production due to entrapment of unreacted residue in the product.	Assets (MPPD)	4	2	5	Level indicator and low level alarm in the reactor.	1	4	Monitor the level indicator of the vessels and ensure the liquid level is maintained within normal range.	1	4
2 5	Low level	Blockage in the gas product stream of the reactor or decrease in the reactor feed flow.	1. Reduced liquid level in the column. 2. Increased gas hold up volume. 3. If the liquid level drops below the height of the feed pipe, gas can potentially backflow into the feed pipe	Assets (MPPD)	4	2	5	Level indicator and low level alarm in the reactor.	1	4	Monitor the level indicator of the vessels and ensure the liquid level is maintained within normal range.	1	4

G.3 DOW FIRE AND EXPLOSION INDECIES

The fire and explosion indexes were evaluated using the Dow's Fire and Explosion Index Hazard Classification Guide, from the American Institute of Chemical Engineers, 1994, to give a comprehensive and realistic potential process unit loss in the case of fire or explosion. For this assessment the fire and explosion indexes based on the material factor using table G.3 as a guide, general process hazards factor and special process hazards factor for each process unit.

After the FEI analysis, the equipment placement on the plot plan was made based on the area of exposure where the equipment could be exposed to a fire or fuel-air explosion. Equipment such as reactors trains (R-01A to R-04B), furnace (F-01 and F-02), and heat exchangers (C-01B and E-2) have a large area of exposure and cover the area of others equipment.

Table G3: Material Factor Determination Guide.

MATERIAL FACTOR DETERMINATION GUIDE						
	Flammability Rankin g	Instability Ranking				
Liquids & Gases Flammability or Combustibility	NFPA 325M or 49	0	1	2	3	4
Non-combustible ²	0	1	14	24	29	40
F.P. > 200 F (> 93.3 C)	1	4	14	24	29	40
F.P. > 100 F (> 37.8 C) 200 F (< 93.3 C)	2	10	14	24	29	40
F.P. 73 F (< 22.8 C) < 100 °F (< 37.8 C) or F.P. < 73 F (< 22.8 C) & BP. 100 F (< 37.8 C)	3	16	16	24	29	40
F.P. < 73 F (< 22.8 C) & BP. < 100 F (< 37.8 C)	4	21	21	24	29	40
Combustible Dust or Mist						
St-1 (K _d = 200 bar m/sec)		16	16	24	29	40
St-2 (K _d = 201-300 bar m/sec)		21	21	24	29	40
St-3 (K _d > 300 bar m/sec)		24	24	24	29	40
Combustible Solids						
Dense > 40 mm thick	1	4	14	24	29	40
Open < 40 mm thick	2	10	14	24	29	40
Foam, fiber, powder, etc.	3	16	16	24	29	40

F.P. = Flash Point, closed cup B.P. = Boiling Point at Standard Temperatures and Pressure (STP)

G.3.1 F&EI Sheets

Location: Edmonton, AB	Business Group: TR Solution			
Prepared by: Naira	Date: March, 2020			
Process Unit: Bitumen Feed Pump				
Materials in Unit: Aromatic Hydrocarbons				
Basis Material for Material Factor: Hydrocarbons				
Material Factor	21.00			
1. General Process Hazards	Penalty Factor Range	Penalty Factor Used		
Base Factor	1.00	1.00		
A. Exothermic Reaction	0.3 - 1.25	0.00		
B. Endothermic Reaction	0.2 - 0.4	0.00		
C. Material Handling and Transfer	0.25 - 0.80	0.85		
D. Enclosed or Indoor Process or storage Units handling Flammable materials	0.00- 0.90	0.00		
E. Ease of Access for Emergency Responders	0.00- 0.35	0.20		
F. Drainage and Spill Control	0.00- 0.50	0.15		
General Process Hazards Factor (F1)	2.20			
2. Special Process Hazards	Penalty Factor Range	Penalty Factor Used		
Base Factor	1.00	1.00		
A. Toxicity of the material handled.	0.00- 0.80	0.20		
B. Process or Storage operates at vacuum	0.50	0.00		
C. Operation in or near the flammable range	0.00 - 0.8	0.30		
D. Dust Explosion	0.00 - 2.0	0.00		
E. Pressure Penalty		0.39		
F. Low Temperature Operation	0.00- 0.30	0.00		
G. Quality of Flammable/ Unstable Material				
1. Combustible and Flammable materials in Process		0.00		
2. Liquids or gases in Storage		0.00		
3. Solids in Storage or Process		0.00		
H. Corrosion and Erosion	0.00 -0.75	0.20		
I. Leakage, Joints, packing, flexible joints	0.00- 1.50	0.30		
J. Use of Fired Equipment		0.00		
K. Hot Oil Heat Exchange Equipment	0.00- 1.15	0.00		
L. Rotating Equipment	0.00- 0.50	0.00		
Special Process Hazards Factor (F2)	2.39			
Process Unit Hazard Factor F3= (F1 x F2)	5.26			

Fire and Explosion Index F&EI= (F3 x MF)		110.53
Location: Edmonton, AB	Business Group: TR Solution	
Prepared by: Naira	Date: March, 2020	
Process Unit: Bitumen Feed Furnace F-01		
Materials in Unit: Aromatic Hydrocarbons		
Basis Material for Material Factor: Hydrocarbons		
Material Factor		24.00
1. General Process Hazards	Penalty Factor Range	Penalty Factor Used
Base Factor	1.00	1.00
A. Exothermic Reaction	0.3 - 1.25	0.00
B. Endothermic Reaction	0.2 - 0.4	0.00
C. Material Handling and Transfer	0.25 - 0.80	0.80
D. Enclosed or Indoor Process or storage Units handling Flammable materials	0.00- 0.90	0.00
E. Ease of Access for Emergency Responders	0.00- 0.35	0.20
F. Drainage and Spill Control	0.00- 0.50	0.25
General Process Hazards Factor (F1)		2.40
2. Special Process Hazards	Penalty Factor Range	Penalty Factor Used
Base Factor	1.00	1.00
A. Toxicity of the material handled.	0.00- 0.80	0.20
B. Process or Storage operates at vacuum	0.50	0.00
C. Operation in or near the flammable range	0.00 - 0.8	0.50
D. Dust Explosion	0.00 - 2.0	0.00
E. Pressure Penalty		0.76
F. Low Temperature Operation	0.00- 0.30	0.00
G. Quality of Flammable/ Unstable Material		
1. Combustible and Flammable materials in Process		0.00
2. Liquids or gases in Storage		0.00
3. Solids in Storage or Process		0.00
H. Corrosion and Erosion	0.00 -0.75	0.50
I. Leakage, Joints, packing, flexible joints	0.00- 1.50	0.10
J. Use of Fired Equipment		0.00
K. Hot Oil Heat Exchange Equipment	0.00- 1.15	0.30
L. Rotating Equipment	0.00- 0.50	0.00
Special Process Hazards Factor (F2)		3.26
Process Unit Hazard Factor F3= (F1 x F2)		7.82

Fire and Explosion Index F&EI= (F3 x MF)		187.71
Location: Edmonton, AB	Business Group: TR Solution	
Prepared by: Naira	Date: March, 2020	
Process Unit: Hydrogen Furnace F-02		
Materials in Unit: Hydrogen		
Basis Material for Material Factor: Hydrogen		
Material Factor		21.00
1. General Process Hazards	Penalty Factor Range	Penalty Factor Used
Base Factor	1.00	1.00
A. Exothermic Reaction	0.3 - 1.25	0.00
B. Endothermic Reaction	0.2 - 0.4	0.00
C. Material Handling and Transfer	0.25 - 0.80	0.85
D. Enclosed or Indoor Process or storage Units handling Flammable materials	0.00- 0.90	0.00
E. Ease of Access for Emergency Responders	0.00- 0.35	0.20
F. Drainage and Spill Control	0.00- 0.50	0.20
General Process Hazards Factor (F1)		2.55
2. Special Process Hazards	Penalty Factor Range	Penalty Factor Used
Base Factor	1.00	1.00
A. Toxicity of the material handled.	0.00- 0.80	0.20
B. Process or Storage operates at vacuum	0.50	0.00
C. Operation in or near the flammable range	0.00 - 0.8	0.50
D. Dust Explosion	0.00 - 2.0	0.00
E. Pressure Penalty		0.76
F. Low Temperature Operation	0.00- 0.30	0.00
G. Quality of Flammable/ Unstable Material		
1. Combustible and Flammable materials in Process		0.00
2. Liquids or gases in Storage		0.00
3. Solids in Storage or Process		0.00
H. Corrosion and Erosion	0.00 -0.75	0.00
I. Leakage, Joints, packing, flexible joints	0.00- 1.50	0.00
J. Use of Fired Equipment		0.00
K. Hot Oil Heat Exchange Equipment	0.00- 1.15	0.00
L. Rotating Equipment	0.00- 0.50	0.00
Special Process Hazards Factor (F2)		2.46

Process Unit Hazard Factor F3= (F1 x F2)		6.27
Fire and Explosion Index F&EI= (F3 x MF)		131.67

Location: Edmonton, AB	Business Group: TR Solution			
Prepared by: Naira	Date: March, 2020			
Process Unit: Overhead Gas Flash Drum D-01				
Materials in Unit: Hydrogen, Methane, Propane, n-Butane, Water, Ammonium, Hydrogen sulfide and Hydrocarbons				
Basis Material for Material Factor: Hydrocarbons				
Material Factor	24.00			
1. General Process Hazards	Penalty Factor Range	Penalty Factor Used		
Base Factor	1.00	1.00		
A. Exothermic Reaction	0.3 - 1.25	0.00		
B. Endothermic Reaction	0.2 - 0.4	0.00		
C. Material Handling and Transfer	0.25 - 0.80	0.80		
D. Enclosed or Indoor Process or storage Units handling Flammable materials	0.00- 0.90	0.00		
E. Ease of Access for Emergency Responders	0.00- 0.35	0.20		
F. Drainage and Spill Control	0.00- 0.50	0.25		
General Process Hazards Factor (F1)		2.25		
2. Special Process Hazards	Penalty Factor Range	Penalty Factor Used		
Base Factor	1.00	1.00		
A. Toxicity of the material handled.	0.00- 0.80	0.20		
B. Process or Storage operates at vacuum	0.50	0.00		
C. Operation in or near the flammable range	0.00 - 0.8	0.50		
D. Dust Explosion	0.00 - 2.0	0.00		
E. Pressure Penalty		0.56		
F. Low Temperature Operation	0.00- 0.30	0.00		
G. Quality of Flammable/ Unstable Material				
1. Combustible and Flammable materials in Process		0.00		
2. Liquids or gases in Storage		0.00		
3. Solids in Storage or Process		0.00		
H. Corrosion and Erosion	0.00 -0.75	0.00		
I. Leakage, Joints, packing, flexible joints	0.00- 1.50	0.00		
J. Use of Fired Equipment		0.00		
K. Hot Oil Heat Exchange Equipment	0.00- 1.15	0.00		

L. Rotating Equipment	0.00- 0.50	0.00
Special Process Hazards Factor (F2)		1.76
Process Unit Hazard Factor F3= (F1 x F2)		3.96
Fire and Explosion Index F&EI= (F3 x MF)		95.13
Location: Edmonton, AB	Business Group: TR Solution	
Prepared by: Naira	Date: March, 2020	
Process Unit: Overhead Gas Flash Drum D-02		
Materials in Unit: Hydrogen, Methane, Propane, n-Butane, Water, Ammonium, Hydrogen sulfide and Hydrocarbons		
Basis Material for Material Factor: Hydrocarbons		
Material Factor	24.00	
1. General Process Hazards	Penalty Factor Range	Penalty Factor Used
Base Factor	1.00	1.00
A. Exothermic Reaction	0.3 - 1.25	0.00
B. Endothermic Reaction	0.2 - 0.4	0.00
C. Material Handling and Transfer	0.25 - 0.80	0.80
D. Enclosed or Indoor Process or storage Units handling Flammable materials	0.00- 0.90	0.00
E. Ease of Access for Emergency Responders	0.00- 0.35	0.20
F. Drainage and Spill Control	0.00- 0.50	0.25
General Process Hazards Factor (F1)		2.25
2. Special Process Hazards	Penalty Factor Range	Penalty Factor Used
Base Factor	1.00	1.00
A. Toxicity of the material handled.	0.00- 0.80	0.20
B. Process or Storage operates at vacuum	0.50	0.00
C. Operation in or near the flammable range	0.00 - 0.8	0.50
D. Dust Explosion	0.00 - 2.0	0.00
E. Pressure Penalty		0.56
F. Low Temperature Operation	0.00- 0.30	0.00
G. Quality of Flammable/ Unstable Material		
1. Combustible and Flammable materials in Process		0.00
2. Liquids or gases in Storage		0.00
3. Solids in Storage or Process		0.00
H. Corrosion and Erosion	0.00 -0.75	0.00
I. Leakage, Joints, packing, flexible joints	0.00- 1.50	0.00
J. Use of Fired Equipment		0.00

K. Hot Oil Heat Exchange Equipment	0.00- 1.15	0.00
L. Rotating Equipment	0.00- 0.50	0.00
Special Process Hazards Factor (F2)		1.76
Process Unit Hazard Factor F3= (F1 x F2)		3.96
Fire and Explosion Index F&EI= (F3 x MF)		95.13
Location: Edmonton, AB	Business Group: TR Solution	
Prepared by: Naira	Date: March, 2020	
Process Unit: Three phase separator D-03		
Materials in Unit: Hydrogen, Methane, Propane, n-Butane, Water, Ammonium, Hydrogen sulfide and Hydrocarbons		
Basis Material for Material Factor: Hydrocarbons		
Material Factor	21.00	
1. General Process Hazards	Penalty Factor Range	Penalty Factor Used
Base Factor	1.00	1.00
A. Exothermic Reaction	0.3 - 1.25	0.00
B. Endothermic Reaction	0.2 - 0.4	0.00
C. Material Handling and Transfer	0.25 - 0.80	0.80
D. Enclosed or Indoor Process or storage Units handling Flammable materials	0.00- 0.90	0.00
E. Ease of Access for Emergency Responders	0.00- 0.35	0.20
F. Drainage and Spill Control	0.00- 0.50	0.25
General Process Hazards Factor (F1)		2.25
2. Special Process Hazards	Penalty Factor Range	Penalty Factor Used
Base Factor	1.00	1.00
A. Toxicity of the material handled.	0.00- 0.80	0.20
B. Process or Storage operates at vacuum	0.50	0.00
C. Operation in or near the flammable range	0.00 - 0.8	0.80
D. Dust Explosion	0.00 - 2.0	0.00
E. Pressure Penalty		0.65
F. Low Temperature Operation	0.00- 0.30	0.00
G. Quality of Flammable/ Unstable Material		
1. Combustible and Flammable materials in Process		0.00
2. Liquids or gases in Storage		0.00
3. Solids in Storage or Process		0.00
M. Corrosion and Erosion	0.00 -0.75	0.00
N. Leakage, Joints, packing, flexible joints	0.00- 1.50	0.00

O. Use of Fired Equipment		0.00
P. Hot Oil Heat Exchange Equipment	0.00- 1.15	0.00
Q. Rotating Equipment	0.00- 0.50	0.00
Special Process Hazards Factor (F2)		2.65
Process Unit Hazard Factor F3= (F1 x F2)		5.96
Fire and Explosion Index F&EI= (F3 x MF)		125.17
Location: Edmonton, AB	Business Group: TR Solution	
Prepared by: Naira	Date: March, 2020	
Process Unit: Slurry Reactor R-01A		
Materials in Unit: Hydrogen, Aromatic Hydrocarbons, Iron (III) Sulfate		
Basis Material for Material Factor: Hydrocarbons		
Material Factor	24.00	
1. General Process Hazards	Penalty Factor Range	Penalty Factor Used
Base Factor	1.00	1.00
A. Exothermic Reaction	0.3 - 1.25	0.00
B. Endothermic Reaction	0.2 - 0.4	0.20
C. Material Handling and Transfer	0.25 - 0.80	0.50
D. Enclosed or Indoor Process or storage Units handling Flammable materials	0.00- 0.90	0.00
E. Ease of Access for Emergency Responders	0.00- 0.35	0.20
F. Drainage and Spill Control	0.00- 0.50	0.25
General Process Hazards Factor (F1)		2.10
2. Special Process Hazards	Penalty Factor Range	Penalty Factor Used
Base Factor	1.00	1.00
A. Toxicity of the material handled.	0.00- 0.80	0.20
B. Process or Storage operates at vacuum	0.50	0.00
C. Operation in or near the flammable range	0.00 - 0.8	0.50
D. Dust Explosion	0.00 - 2.0	0.00
E. Pressure Penalty		0.76
F. Low Temperature Operation	0.00- 0.30	0.00
G. Quality of Flammable/ Unstable Material		
1. Combustible and Flammable materials in Process		0.00
2. Liquids or gases in Storage		0.00
3. Solids in Storage or Process		0.00
H. Corrosion and Erosion	0.00 -0.75	0.20
I. Leakage, Joints, packing, flexible joints	0.00- 1.50	0.10

J. Use of Fired Equipment		0.00
K. Hot Oil Heat Exchange Equipment	0.00- 1.15	0.75
L. Rotating Equipment	0.00- 0.50	0.00
Special Process Hazards Factor (F2)		2.76
Process Unit Hazard Factor F3= (F1 x F2)		5.80
Fire and Explosion Index F&EI= (F3 x MF)		139.03

Location: Edmonton, AB	Business Group: TR Solution			
Prepared by: Naira	Date: March, 2020			
Process Unit: Airfin Cooler C-01B				
Materials in Unit: Hydrogen, Methane, Propane, n-Butane, Water, Ammonium, Hydrogen sulfide				
Basis Material for Material Factor: Hydrocarbons				
Material Factor	21.00			
1. General Process Hazards	Penalty Factor Range	Penalty Factor Used		
Base Factor	1.00	1.00		
A. Exothermic Reaction	0.3 - 1.25	0.00		
B. Endothermic Reaction	0.2 - 0.4	0.00		
C. Material Handling and Transfer	0.25 - 0.80	0.80		
D. Enclosed or Indoor Process or storage Units handling Flammable materials	0.00- 0.90	0.00		
E. Ease of Access for Emergency Responders	0.00- 0.35	0.20		
F. Drainage and Spill Control	0.00- 0.50	0.25		
General Process Hazards Factor (F1)		2.25		
2. Special Process Hazards	Penalty Factor Range	Penalty Factor Used		
Base Factor	1.00	1.00		
A. Toxicity of the material handled.	0.00- 0.80	0.20		
B. Process or Storage operates at vacuum	0.50	0.00		
C. Operation in or near the flammable range	0.00 - 0.8	0.80		
D. Dust Explosion	0.00 - 2.0	0.00		
E. Pressure Penalty		0.76		
F. Low Temperature Operation	0.00- 0.30	0.00		
G. Quality of Flammable/ Unstable Material				
1. Combustible and Flammable materials in Process		0.00		
2. Liquids or gases in Storage		0.00		
3. Solids in Storage or Process		0.00		

H. Corrosion and Erosion	0.00 -0.75	0.10
I. Leakage, Joints, packing, flexible joints	0.00- 1.50	0.10
J. Use of Fired Equipment		0.00
K. Hot Oil Heat Exchange Equipment	0.00- 1.15	0.00
L. Rotating Equipment	0.00- 0.50	0.00
Special Process Hazards Factor (F2)		2.96
Process Unit Hazard Factor F3= (F1 x F2)		6.66
Fire and Explosion Index F&EI= (F3 x MF)		139.76

Location: Edmonton, AB	Business Group: TR Solution			
Prepared by: Naira	Date: March, 2020			
Process Unit: Reactor Overhead Cooler E-02				
Materials in Unit: Hydrogen, Methane, Propane, n-Butane, Water, Ammonium, Hydrogen sulfide				
Basis Material for Material Factor: Hydrocarbons				
Material Factor	24.00			
1. General Process Hazards	Penalty Factor Range	Penalty Factor Used		
Base Factor	1.00	1.00		
A. Exothermic Reaction	0.3 - 1.25	0.00		
B. Endothermic Reaction	0.2 - 0.4	0.00		
C. Material Handling and Transfer	0.25 - 0.80	0.80		
D. Enclosed or Indoor Process or storage Units handling Flammable materials	0.00- 0.90	0.00		
E. Ease of Access for Emergency Responders	0.00- 0.35	0.20		
F. Drainage and Spill Control	0.00- 0.50	0.25		
General Process Hazards Factor (F1)		2.25		
2. Special Process Hazards	Penalty Factor Range	Penalty Factor Used		
Base Factor	1.00	1.00		
A. Toxicity of the material handled.	0.00- 0.80	0.20		
B. Process or Storage operates at vacuum	0.50	0.00		
C. Operation in or near the flammable range	0.00 - 0.8	0.30		
D. Dust Explosion	0.00 - 2.0	0.00		
E. Pressure Penalty		0.99		
F. Low Temperature Operation	0.00- 0.30	0.00		
G. Quality of Flammable/ Unstable Material				

1. Combustible and Flammable materials in Process		0.00
2. Liquids or gases in Storage		0.00
3. Solids in Storage or Process		0.00
H. Corrosion and Erosion	0.00 -0.75	0.10
I. Leakage, Joints, packing, flexible joints	0.00- 1.50	0.10
J. Use of Fired Equipment		0.00
K. Hot Oil Heat Exchange Equipment	0.00- 1.15	0.00
L. Rotating Equipment	0.00- 0.50	0.00
Special Process Hazards Factor (F2)		2.69
Process Unit Hazard Factor F3= (F1 x F2)		6.05
Fire and Explosion Index F&EI= (F3 x MF)		145.01

Table G4. Process units F&EI and area of exposure summary.

Equipment	Fire & Explosion Index	Radius of Exposure (m)	Area of Exposure (m ²)
P-01	111	27.4	2359
F-01	188	48.8	7482
F-02	132	33.5	3526
D-01	95	24.4	1870
D-02	95	24.4	1870
D-03	125	32.0	3217
R-01	139	35.7	4004
C-01B	140	36.6	4208
E-02	145	37.8	4489

G.4 DOW CHEMICAL EXPOSURE INDEX

The chemical exposure indexes were evaluated using the Dow's Chemical Exposure Index Guide, from the American Institute of Chemical Engineers, 1994, to estimate the airborne quantity released and rate the potential health hazard to neighbor plants and communities in the event of chemical release incidents. For this assessment the chemical exposure indexes based on Emergency Response Planning Guidelines (ERPG) and Dow Emergency Exposure Planning Guidelines (EEPG), using table G5 as a guide. Then the airborne quantity for gas and liquid releases are estimated, following by the CEI and hazard distance calculation.

Table G5: ERPGs and Physical Properties.

	Molecular Weight	Boiling Point (°C)	ERPG-1 (mg/m3)	ERPG-2 (ppm)	ERPG-2 (mg/m3)	ERPG-3 (ppm)	ERPG-3 (mg/m3)	Vapour Pressure (kPa)	Liquid Density (kg/m3)	Heat Capacity (J/kg·°C)	Heat of Vaporization (J/kg)	Latent Heat (J/kg)	Ratio of Cp/Hv (1/°C)
Acetone cyanohydrin *	85.1	-	-	-	35	10	-	-	-	-	-	-	-
Acrolein	56.1	52.5	-	0.1	1	0.5	7	3	36	835.1	2.00E+03	5.37E+05	0.0037
Acrylic acid	72.1	141.4	6	2	147	50	2210	750	0.5	1046.7	1.80E+03	6.95E+05	0.0026
Acrylonitrile *	53.1	77.2	22	10	76	35	163	75	14	778.7	2.06E+03	6.31E+05	0.0033
Allyl chloride	76.5	44.8	9	3	125	40	939	300	48.6	932.1	1.53E+03	3.73E+05	0.0041
Ammonia	17	-33.4	17	25	104	150	522	750	1013.1	602.2	4.74E+03	1.16E+06	0.0041
Benzene	78.1	80.1	-	-	-	-	-	-	12.7	874.5	1.74E+03	4.30E+05	0.004
Bromine	159.8	58.7	1	0.2	7	1	33	5	28.5	3106.4	4.65E+02	1.96E+05	0.0024
Butadiene	54.1	-4.4	22	10	442	200	11060	5000	-	-	-	-	-
n-butyl acrylate	128.2	-	0.26	0.05	131	25	1310	250	0.7	894.5	1.82E+03	3.61E+05	0.005
n-butylisocyanate	99.1	-	0.04	0.01	0.2	0.05	4	1	2.6	886.6	1.60E+03	4.07E+05	0.0039
Carbon disulfide	76.1	46.3	3	1	156	50	1557	500	48.3	1256.3	1.09E+03	3.61E+05	0.003
Carbon monoxide	28	-191.5	-	200	-	350	-	500	-	-	-	-	-
Carbon tetrachloride	153.8	76.8	126	20	629	100	4718	750	15.2	1582.9	8.24E+02	2.11E+05	0.0039
Chlorine	70.9	-34	3	1	9	3	58	20	779.4	1400.3	9.85E+02	2.47E+05	0.004
Chlorine trifluoride	92.5	-	0.38	0.1	4	1	38	10	-	-	-	-	-
Chloroacetyl chloride	112.9	106	0.25	0.05	2.5	0.5	23	5	3.3	1410.8	1.17E+03	3.62E+05	0.0032
Chloroform *	119.4	61.7	-	-	244	50	24413	5000	26.3	1480.8	9.54E+02	2.61E+05	0.0037
Chloropicrin	164.4	112	1	0.1	2	0.3	10	1.5	3.4	1646	8.55E+02	2.36E+05	0.0036
Chlorosulfonic acid	116.5	-	2	0.4	10	2.1	30	6.3	-	-	-	-	-
Chlorotrifluoroethylene	116.5	-28.2	95	20	476	100	1429	300	641.9	1292	1.20E+03	1.46E+05	0.0082
Crotonaldehyde	70.1	102.4	6	2	29	10	143	50	4.8	849.4	2.00E+03	5.45E+05	0.0037
Diketene	82.1	-	3	1	17	5	168	50	-	-	-	-	-
Dimethylamine	45.1	6.9	2	1	184	100	922	500	205.9	650.4	2.77E+03	5.61E+05	0.0049
Epichlorohydrin	92.5	116.4	8	2	76	20	378	100	2.3	1175.1	1.44E+03	4.51E+05	0.0032
Ethyl chloride *	64.5	12.3	-	-	13192	5000	-	-	160.3	893	1.62E+03	3.74E+05	0.0043
Ethylene dichloride *	99	83.5	-	-	405	100	-	-	10.6	1247.1	1.31E+03	3.54E+05	0.0037
Ethylene oxide *	44.1	10.5	-	-	90	50	901	500	174.4	867.7	2.07E+03	5.61E+05	0.0037
Formaldehyde	30	-	1	1	12	10	31	25	515.6	728.7	2.55E+03	7.13E+05	0.0036
Hexachlorobutadiene	260.8	-	32	3	107	10	320	30	0	1673.9	6.83E+02	2.14E+05	0.0032
Hexafluoroacetone	166	-	-	-	7	1	339	50	-	-	-	-	-
Hydrogen bromide *	80.9	-66.7	10	3	66	20	248	75	2359.7	1760.3	1.02E+03	1.59E+05	0.0064
Hydrogen chloride	36.5	-85	4	3	30	20	224	150	4812.5	804.9	3.40E+03	2.36E+05	0.0144
Hydrogen cyanide	27	25.7	-	-	11	10	28	25	-	-	-	-	-
Hydrogen fluoride	20	19.6	4	5	16	20	41	50	123	983.3	#####	3.72E+05	-0.0019
Hydrogen sulfide	34.1	-60.3	0.14	0.1	42	30	139	100	2022.4	767.2	2.33E+03	4.11E+05	0.0057
Isobutyronitrile	69.1	-	28	10	141	50	565	200	-	-	-	-	-
2-isocyanoethyl methacrylate	155.2	-	-	-	1	0.1	6	1	0	1095.9	1.83E+03	3.58E+05	0.0051
Methacrylonitrile *	67.1	90.3	-	-	27	10	-	-	9.5	795.5	1.99E+03	4.89E+05	0.0041
Methanol	32	64.5	262	200	1310	1000	6551	5000	17	786.7	2.54E+03	1.20E+06	0.0021
Methyl chloride	50.5	-24.1	-	-	826	400	2065	1000	577.3	916.9	1.62E+03	3.76E+05	0.0043
Methyl iodide	141.9	-	145	25	290	50	726	125	-	-	-	-	-
Methyl isocyanate	57.1	-	0.058	0.025	1	0.5	12	5	62	947.8	1.44E+03	4.90E+05	0.0029
Methyl mercaptan	48.1	6	0.01	0.005	49	25	197	100	202.2	857.7	1.89E+03	4.87E+05	0.0039
Monomethylamine	31.1	-6.3	13	10	127	100	635	500	349.1	655.9	3.10E+03	7.77E+05	0.004
Perfluoroisobutylene	218.1	-	-	-	1	0.1	3	0.3	-	-	-	-	-
Phenol	94.1	181.9	38	10	192	50	770	200	0.1	1070.6	2.05E+03	6.22E+05	0.0033
Phosgene	98.9	7.5	-	-	1	0.2	4	1	190.3	1361.7	1.03E+03	2.37E+05	0.0043
Phosphorous pentoxide	141.9	-	5	1	25	4	100	17	-	-	-	-	-
Propylene oxide *	58.1	34.2	119	50	594	250	1782	750	71.9	823.7	2.09E+03	4.80E+05	0.0043
Styrene	104.2	145.2	213	50	1065	250	4259	1000	0.8	902	1.75E+03	4.22E+05	0.0041
Sulfur dioxide	64.1	-10	1	0.3	8	3	39	15	393.6	1352.4	1.41E+03	3.57E+05	0.0039
Sulfuric acid (Sulfur trioxide)	98.1	44.4	2	0.5	10	2.5	30	7.5	-	-	-	-	-
Sulfuryl fluoride *	102.1	-	-	-	417	100	1252	300	1748	1321	1.50E+03	1.28E+05	0.0117
Tetrafluoroethylene	100	-	818	200	4090	1000	40902	10000	3249.7	880.9	3.48E+03	6.31E+04	0.0551
Titanium tetrachloride	189.7	-	5	1	20	3	100	13	1.4	1715.8	7.43E+02	2.16E+05	0.0034
Toluene diisocyanate *	174.2	252.9	-	-	1	0.2	-	-	0	1210.8	1.52E+03	3.92E+05	0.0039
Trimethylamine	59.1	2.9	-	0.1	242	100	1209	500	221.5	625.5	2.29E+03	3.68E+05	0.0062
Vinyl acetate	86.1	72.8	18	5	264	75	1760	500	15.3	925.4	1.76E+03	4.03E+05	0.0044
Vinyl chloride *	62.5	-13.8	-	-	2556	1000	12781	5000	395.7	903.2	1.55E+03	3.20E+05	0.0048
Vinylidene chloride *	96.9	31.7	-	-	198	50	793	200	79.7	1204	1.16E+03	2.74E+05	0.0042

ESTIMATING THE AIRBORNE QUANTITY FOR GAS RELEASES

The following equations, based on the sonic gas flow rate equation, are used to estimate the airborne quantity for a gas release.

SI Units

$$\text{Airborne Quantity (AQ)} = 4.751 \times 10^{-6} D^2 P_a \sqrt{\frac{MW}{T + 273}} \quad \text{(kg/sec)} \quad (\text{Equation 1A})$$

where

P_a = absolute pressure = ($P_g + 101.35$)

P_g = gauge pressure (kPa gauge)

MW = molecular weight of the material

T = temperature (°C)

D = diameter of the hole (millimeters)

ESTIMATING THE AIRBORNE QUANTITY FOR LIQUID RELEASES

The following steps describe a simplified procedure for estimating the airborne quantity for liquid releases.

Step 1: Determine the liquid flow rate being released.

The liquid release rate (L) is given by the following equations:

These equations assume that release from all scenarios will continue for at least five minutes before releases can be stopped. If a five minute release would exceed the total inventory, the release rate is calculated by dividing the total inventory by five minutes.

SI Units

$$L = 9.44 \times 10^{-7} D^2 \rho_1 \sqrt{\frac{1000 P_g}{\rho_1} + 9.8 \Delta h} \quad \text{(kg/sec)} \quad (\text{Equation 2A})$$

where

P_g = gauge pressure (kPa gauge)

(Note: for a tank open to the atmosphere $P_g = 0$)

ρ_1 = density of the liquid at operating temperature (kg/m³)

Δh = height of the liquid above the release point (meters)

D = diameter of the hole (millimeters)

CEI AND HAZARD DISTANCE CALCULATION

Chemical Exposure Index

All CEI calculations assume a windspeed of 5 m/sec (11.2 miles/hour) and neutral weather conditions.

The Chemical Exposure Index (CEI) is given by:

SI Units

$$\text{CEI} = 655.1 \sqrt{\frac{\text{AQ}}{\text{ERPG-2}}} \quad (\text{Equation 10A})$$

where

AQ = airborne quantity (kg/sec)

ERPG-2 = value (mg/m³)

Hazard Distance

The Hazard Distance (HD) is the distance to the ERPG-1, -2 or -3 concentration and is derived from the following equation:

SI Units

$$\text{HD} = 6551 \sqrt{\frac{\text{AQ}}{\text{ERPG}}} \quad \{\text{meters}\} \quad (\text{Equation 11A})$$

where

AQ = airborne quantity (kg/sec)

ERPG = ERPG-1, ERPG-2 or ERPG-3 (mg/m³)

G.4.1 CEI Sheets

Table G6. Hydrogen sulfide vapor release for stream 30.

Piping release or vessel nozzle release	
Quantity of gas available for release kg	25914.5
Temperature of released material deg C	75
Molecular Weight	34.1
Boiling Point C	-60.3
Liquid Density kg/M3 at storage temp C	669.1
Liquid Density kg/M3 at BP	767.2
Gas Density kg/M3 at 25C	1.36
Heat Capacity C _p Joules/kg-deg C	2.33E+03
Latent Heat of Vaporisation H _v Joules/Kg	4.11E+05
Ratio of C _p /H _v	0.0057
Absolute Pressure (P _a) kPa	25100
Gauge Pressure in Process (P _g) kPa	25000
Temperature C	75
Diameter of hole mm	76.2
Height of leak above grade (Metres)	6
ERPG2 Mg/M3	42
ERPG3 Mg/M3	139
Dike area M2	0.018
Gas Releases	
Airborne Gas Release (continuous assuming large inventory) kg/sec	96.33
Airborne Gas Release (inventory exhausted within 5 minutes) kg/sec	86.38
CEI	939
Distance travel by ERPG2 concentration (Meters)	9395
Distance travel by ERPG3 concentration (Meters)	5164
Maximum Release Duration (Minutes)	5

Table G7: Hydrogen sulfide vapor release for stream 52.

Piping release	
Quantity of liquid available for release kg	
Quantity of gas available for release kg	489.2
Temperature of released material deg C	75.9
Molecular Weight	34.1
Boiling Point C	-60.3
Liquid Density kg/M3 at storage temp C	669.1
Liquid Density kg/M3 at BP	104.2
Gas Density kg/M3 at 25C	1.36
Heat Capacity C _p Joules/kg-deg C	2.33E+03
Latent Heat of Vaporisation H _v Joules/Kg	4.11E+05
Ratio of Cp/Hv	0.0057
Absolute Pressure (P _a) kPa	190
Gauge Pressure in Process (P _g) kPa	90
Temperature C	75.9
Diameter of hole mm	76.2
Height of leak above grade (Metres)	10
ERPG2 Mg/M3	42
ERPG3 Mg/M3	139
Dike area M2	0.018
Gas Releases	
Airborne Gas Release (continuous assuming large inventory) kg/sec	0.73
Airborne Gas Release (inventory exhausted within 5 minutes) kg/sec	0.73
CEI	86
Distance travel by ERPG2 concentration (Meters)	863
Distance travel by ERPG3 concentration (Meters)	474
Maximum Release Duration (Minutes)	11

Table G8. Vapor ammonia release for stream 52.

Piping release	
Quantity of gas available for release kg	2614.8
Temperature of released material deg C	75.9
Molecular Weight	17.031
Boiling Point C	-33.4
Vapour Pressure t 25 C kPa	1013.1
Liquid Density kg/M3 at storage temp C	602.2
Liquid Density kg/M3 at BP	704
Gas Density kg/M3 at 25C	0.771
Heat Capacity C _p Joules/kg-deg C	4.74E+03
Latent Heat of Vaporisation H _v Joules/Kg	1.16E+06
Ratio of Cp/Hv	0.0041
Absolute Pressure (P _a) kPa	190
Gauge Pressure in Process (P _g) kPa	90
Temperature C	75.9
Diameter of hole mm	76.2
Height of leak above grade (Metres)	10
ERPG2 Mg/M3	104
ERPG3 Mg/M3	522
Dike area M2	0.018
Gas Releases	
Airborne Gas Release (continuous assuming large inventory) kg/sec	0.51
Airborne Gas Release (inventory exhausted within 5 minutes) kg/sec	0.51
CEI	46
Distance travel by ERPG2 concentration (Meters)	461
Distance travel by ERPG3 concentration (Meters)	206
Maximum Release Duration (Minutes)	5

Table G9. Liquid ammonia release for stream 30.

Piping release or vessel nozzle release	
Quantity of liquid available for release kg	25914.5
Quantity of gas available for release kg	
Temperature of released material deg C	75
Molecular Weight	17.031
Boiling Point C	-33.4
Vapour Pressure t 25 C kPa	1013.1
Vapour Pressure at Pool Temperature kPa	
Liquid Density kg/M3 at storage temp C	602.2
Liquid Density kg/M3 at BP	704
Gas Density kg/M3 at 25C	0.771
Heat Capacity C _p Joules/kg-deg C	4.74E+03
Latent Heat of Vaporisation H _v Joules/Kg	1.16E+06
Ratio of Cp/Hv	0.0041
Absolute Pressure (P _a) kPa	25100
Gauge Pressure in Process (P _g) kPa	25000
Temperature C	75
Diameter of hole mm	76.2
Height of leak above grade (Metres)	10
ERPG2 Mg/M3	104
ERPG3 Mg/M3	522
Dike area M2	0.018
Liquid Releases	
Liquid release rate L kg/sec	86.38
Release quantity (unlimited) kg	77743.5
Actual Release Quantity (limited by emergency block valve or quantity available) kg	25914.5
Fraction Flashed	0.44
Airborne Quantity produced by the flash kg/sec	86.38
Characteristic pool temperature (Deg C)	-33.4
Total Airborne Quantity (kg/sec)	86.38
Chemical Exposure Index	597
Distance travel by ERPG2 concentration (Meters)	5970
Distance travel by ERPG3 concentration (Meters)	2665
Maximum Release Duration (Minutes)	6

Table G10. Chemical exposure index summary.

Stream	Chemical Released	CEI	Hazard Distance (m)	Maximum Release Duration (min)
30	Vapor H ₂ S	939	9395	5
52	Vapor H ₂ S	86	863	11
52	Vapor Ammonia	46	461	5
30	Liquid Ammonia	597	5970	6

G.5 MATERIAL SAFETY DATA SHEETS

The following sheets of material safety data sheets (SDS) for the chemicals involved in the plant process contain information on potential hazards and how to work safely around them. The SDS for the different chemicals gases and liquids were retrieved from Suncor Energy and Praxair Canada both in compliance with WHMIS 2015.

Ammonia - US

Safety Data Sheet P-4562

This SDS conforms to U.S. Code of Federal Regulations 29 CFR 1910.1200, Hazard Communication.

Date of issue: 01/01/1981 Revision date: 08/28/2019 Supersedes: 03/23/2015

SECTION 1: Product and company identification

1.1. Product identifier

Product form	:	Substance
Substance name	:	Ammonia - US
CAS-No.	:	7664-41-7
Formula	:	NH3

1.2. Relevant identified uses of the substance or mixture and uses advised against

Use of the substance/mixture	:	Industrial use; Use as directed.
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1.3. Details of the supplier of the safety data sheet

Praxair, Inc.
10 Riverview Drive
Danbury, CT 06810-6268 - USA
T 1-800-772-9247 (1-800-PRAIR) - F 1-716-879-2146
www.praxair.com

1.4. Emergency telephone number

Emergency number	:	Onsite Emergency: 1-800-645-4633
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CHEMTREC, 24hr/day 7days/week
— Within USA: 1-800-424-9300, Outside USA: 001-703-527-3887
(collect calls accepted, Contract 17729)

SECTION 2: Hazard identification

2.1. Classification of the substance or mixture

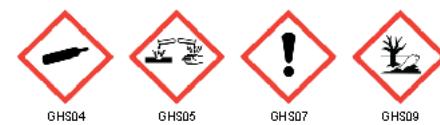
GHS US classification

Flam. Gas 2	H221
Press. Gas (Liq.)	H280
Acute Tox. 4 (Inhalation:gas)	H332
Skin Corr. 1B	H314
STOT SE 3'	H335
Aquatic Acute 1	H400

2.2. Label elements

GHS US labeling

Hazard pictograms (GHS US)



GHS04

GHS05

GHS07

GHS09

Signal word (GHS US)

: Danger

Hazard statements (GHS US)

: H221 - FLAMMABLE GAS
H280 - CONTAINS GAS UNDER PRESSURE; MAY EXPLODE IF HEATED
H314 - CAUSES SEVERE SKIN BURNS AND EYE DAMAGE
H332 - HARMFUL IF INHALED
H400 - VERY TOXIC TO AQUATIC LIFE
CGA-HG01 - MAY CAUSE FROSTBITE.
CGA-HG22 - CORROSIVE TO THE RESPIRATORY TRACT (This statement supercedes H335)

Precautionary statements (GHS US)

: P202 - Do not handle until all safety precautions have been read and understood.
P210 - Keep away from Heat, Open flames, Sparks, Hot surfaces. - No smoking
P260 - Do not breathe gas
P262 - Do not get in eyes, on skin, or on clothing.
P280 - Wear protective gloves, protective clothing, eye protection, face protection.



Butane

Safety Data Sheet E-4572

according to the Hazardous Products Regulation (February 11, 2015);
Date of Issue: 10-15-1979 Revision date: 09-18-2018 Supersedes: 10-15-2016

SECTION 1: Identification

1.1. Product Identifier

Product form	:	Substance
Trade name	:	Butane
Chemical name	:	Bulane
CAS No.	:	106-97-8
Formula	:	C4H10
Other means of identification	:	Methylethylmethane, Diethyl, n-Butane, Butyl hydride.
Product group	:	Core Products

1.2. Recommended use and restrictions on use

Recommended uses and restrictions	:	Industrial use Welding
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1.3. Supplier

Praxair Canada Inc.
1200 – 1 City Centre Drive
Mississauga - Canada L5B 1M2
T 1-905-803-1600 - F 1-905-803-1682
www.praxair.ca

1.4. Emergency telephone number

Emergency number	:	1-800-363-0042
		Call emergency number 24 hours a day only for spills, leaks, fire, exposure, or accidents involving this product.
		For routine information, contact your supplier or Praxair sales representative.

SECTION 2: Hazard identification

2.1. Classification of the substance or mixture

GHS-CA classification

Flam. Gas 1 H220
Liquefied gas H280

2.2. GHS Label elements, including precautionary statements

GHS-CA labelling

Hazard pictograms



Signal word

:

DANGER

Hazard statements

:

EXTREMELY FLAMMABLE GAS
CONTAINS GAS UNDER PRESSURE; MAY EXPLODE IF HEATED
MAY CAUSE FROSTBITE.
MAY FORM EXPLOSIVE MIXTURES WITH AIR.

Precautionary statements

:

Keep away from heat, hot surfaces, sparks, open flames and other ignition sources. No smoking.
Use and store only outdoors or in a well-ventilated area.
Leaking gas fire: Do not extinguish, unless leak can be stopped safely.
In case of leakage, eliminate all ignition sources.
Protect from sunlight when ambient temperature exceeds 52 °C (125°F).
Use a back flow preventive device in the piping.

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Carbon dioxide, Safety Data Sheet E-4574

according to the Hazardous Products Regulation (February 11, 2015)

Date of Issue: 10-15-1979

Revisor date: 08-03-2016

Supersedes: 10-15-2014

SECTION 1: Identification

1.1. Product Identifier

Product form	:	Substance
Name	:	Carbon dioxide,
CAS No	:	124-38-9
Formula	:	CO ₂
Other means of identification	:	Carbon anhydride, Carbonic acid gas, Carbon Dioxide,
Product group	:	Core Products

1.2. Recommended use and restrictions on use

Recommended uses and restrictions	:	Industrial use Medical applications Semiconductor Use as directed
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1.3. Supplier

Praxair Canada Inc.
1200 1 City Centre Drive
Mississauga - Canada L5B 1M2
T 1-905-803-1600 - F 1-905-803-1682
www.praxair.ca

1.4. Emergency telephone number

Emergency number	:	1-800-363-0042 Call emergency number 24 hours a day only for spills, leaks, fire, exposure, or accidents involving this product. For routine information, contact your supplier or Praxair sales representative.
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SECTION 2: Hazard identification

2.1. Classification of the substance or mixture

GHS-CA classification

Simple asphyxiant: I (I30)
Compressed gas H280

2.2. GHS Label elements, including precautionary statements

GHS-CA labelling

Hazard pictograms



GHS02

Signal word

:

WARNING

Hazard statements

:

CONTAINS REFRIGERATED GAS; MAY CAUSE CRYOGENIC BURNS OR INJURY
MAY DISPLACE OXYGEN AND CAUSE RAPID SUFFOCATION
MAY CAUSE FROSTBITE
MAY INCREASE RESPIRATION AND HEART RATE

Precautionary statements

:

Do not handle until all safety precautions have been read and understood
Use and store only outdoors or in a well-ventilated area
Wear cold insulating gloves and either face shield or eye protection
Protect from sunlight when ambient temperature exceeds 52°C (125°F)
Use a back flow preventive device in the piping
Close valve after each use and when empty

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Carbon monoxide

Safety Data Sheet E-4576

according to the Hazardous Products Regulation (February 11, 2015)

Date of Issue: 10-15-1979

Revision date: 08-05-2016

Supersedes: 10-15-2013

SECTION 1: Identification

1.1. Product Identifier

Product form	:	Substance
Name	:	Carbon monoxide
CAS No	:	630-08-0
Formula	:	CO
Other means of identification	:	Carbon monoxide, compressed Compressed carbon monoxide Carbon oxide (CO)
Product group	:	Core Products

1.2. Recommended use and restrictions on use

Recommended uses and restrictions	:	Industrial use Use as directed
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1.3. Supplier

Praxair Canada Inc.
1200 - 1 City Centre Drive
Mississauga - Canada L5B 1M2
T 1-905-803-1600 - F 1-905-803-1682
www.praxair.ca

1.4. Emergency telephone number

Emergency number	:	1-800-363-0042 Call emergency number 24 hours a day only for spills, leaks, fire, exposure, or accidents involving this product. For routine information, contact your supplier or Praxair sales representative.
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SECTION 2: Hazard identification

2.1. Classification of the substance or mixture

GHS-CA classification

Flam. Gas 1	H120
Compressed gas	H280
Acute Tox. 3 (Inhalation:gas)	H331
Repr. 1A	H360
STOT RF 1	H372

2.2. GHS Label elements, including precautionary statements

GHS-CA labelling

Hazard pictograms



GHS02

GHS04

GHS06

GHS08

Signal word

:

DANGER

Hazard statements

: EXTREMELY FLAMMABLE GAS
CONTAINS GAS UNDER PRESSURE; MAY EXPLODE IF HEATED
TOXIC IF INHALED
May damage fertility or the unborn child
CAUSES DAMAGE TO ORGANS (CENTRAL NERVOUS SYSTEM) THROUGH PROLONGED OR REPEATED EXPOSURE (Inhalation)
MAY FORM EXPLOSIVE MIXTURES WITH AIR

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Ethane

Safety Data Sheet E-4592

according to the Hazardous Products Regulation (February 11, 2015)

Date of Issue: 10-15-1979

Revised date: 08-15-2016

Supersedes: 10-15-2013

SECTION 1: Identification

1.1. Product Identifier

Product form	:	Substance
Name	:	Ethane
CAS No.	:	74-84-0
Formula	:	C ₂ H ₆
Other means of identification	:	Methylmethane, bimethyl, dimethyl, ethyl hydride
Product group	:	Core Products

1.2. Recommended use and restrictions on use

Recommended uses and restrictions	:	Industrial use Use as directed
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1.3. Supplier

Praxair Canada inc.
1200 - 1 City Centre Drive
Mississauga - Canada L5B 1M2
T 1-905-803-1600 - F 1-905-803-1682
www.praxair.ca

1.4. Emergency telephone number

Emergency number	:	1-800-363-0042
		Call emergency number 24 hours a day only for spills, leaks, fire, exposure, or accidents involving this product.
		For routine information, contact your supplier or Praxair sales representative.

SECTION 2: Hazard identification

2.1. Classification of the substance or mixture

GHS-CA classification

Flam. Gas 1 H1220
Compressed gas H280

2.2. GHS Label elements, including precautionary statements

GHS-CA labelling

Hazard pictograms



GHS02

GHS07

Signal word

:

DANGER

Hazard statements

:

EXTREMELY FLAMMABLE GAS
CONTAINS GAS UNDER PRESSURE; MAY EXPLODE IF HEATED
MAY DISPLACE OXYGEN AND CAUSE RAPID SUFFOCATION
MAY CAUSE FROSTBITE
MAY FORM EXPLOSIVE MIXTURES WITH AIR

Precautionary statements

:

- Do not handle until all safety precautions have been read and understood
Keep away from heat, hot surfaces, sparks, open flames and other ignition sources. No smoking
- Use and store only outdoors or in a well-ventilated area
Leaking gas fire: Do not extinguish, unless leak can be stopped safely
In case of leakage, eliminate all ignition sources
Protect from sunlight when ambient temperature exceeds 52°C (125°F)

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Hydrogen sulfide

Safety Data Sheet E-4611

according to the Hazardous Products Regulation (February 11, 2015)

Date of Issue: 10-15-1979

Revised date: 08-10-2016

Supersedes: 10-15-2013

SECTION 1: Identification

1.1. Product Identifier

Product form	:	Substance
Name	:	Hydrogen sulfide
CAS No.	:	7783-06-4
Formula	:	H2S
Other means of identification	:	Hydrogen sulfide
Product group	:	Core Products

1.2. Recommended use and restrictions on use

Recommended uses and restrictions	:	Industrial use Use as directed
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1.3. Supplier

Praxair Canada inc.
1200 - 1 City Centre Drive
Mississauga - Canada L5B 1M2
T 1-905-803-1600 - F 1-905-803-1682
www.praxair.ca

1.4. Emergency telephone number

Emergency number	:	1-800-363-0042
	:	Call emergency number 24 hours a day only for spills, leaks, fire, exposure, or accidents involving this product.
	:	For routine information, contact your supplier or Praxair sales representative.

SECTION 2: Hazard identification

2.1. Classification of the substance or mixture

GHS-CA classification

Flam. Gas 1	H120
Liquefied gas	H280
Acute Tox. 2 (Inhalation: gas)	H330
STOT SE 3	H335

2.2. GHS Label elements, including precautionary statements

GHS-CA labelling

Hazard pictograms



Signal word

:

DANGER

Hazard statements

:

EXTREMELY FLAMMABLE GAS
CONTAINS GAS UNDER PRESSURE; MAY EXPLODE IF HEATED
FATAL IF INHALED
MAY CAUSE RESPIRATORY IRRITATION
MAY FORM EXPLOSIVE MIXTURES WITH AIR
SYMPTOMS MAY BE DELAYED
EXTENDED EXPOSURE TO GAS REDUCES THE ABILITY TO SMELL SULFIDES

Precautionary statements

:

Do not handle until all safety precautions have been read and understood
Keep away from heat, hot surfaces, sparks, open flames and other ignition sources. No smoking

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Hydrogen, compressed

Safety Data Sheet P-4604

This SDS conforms to U.S. Code of Federal Regulations 29 CFR 1910.1200, Hazard Communication.

Date of issue: 01/01/1980 Revision date: 10/17/2016 Supersedes: 06/03/2015

SECTION 1. Product and company identification

1.1 Product identifier

Product form	:	Substance
Name	:	Hydrogen, compressed
CAS No	:	1333-74-0
Formula	:	H2
Other means of identification	:	Dihydrogen, parahydrogen, refrigerant gas R702, water gas

1.2 Relevant identified uses of the substance or mixture and uses advised against

Use of the substance/mixture	:	Industrial use. Use as directed.
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1.3 Details of the supplier of the safety data sheet

Praxair, Inc.
10 Riverview Drive
Danbury, CT 06810-6268 - USA
T 1-800-772-9247 (1-800-PRAXAIR) - F 1-716-879-2146
www.praxair.com

1.4 Emergency telephone number

Emergency number	:	Onsite Emergency: 1-800-645-4633
------------------	---	----------------------------------

CHEMTREC, 24hr/day 7days/week
— Within USA: 1-800-424-9300, Outside USA: 001-703-527-3887
(collect calls accepted, Contract 17729)

SECTION 2. Hazard identification

2.1 Classification of the substance or mixture

GHS-US classification

Flam. Gas 1 H220
Compressed gas H280

2.2 Label elements

GHS-US labeling

Hazard pictograms (GHS-US)



GHS02

GHS04

Signal word (GHS-US)

:

DANGER

Hazard statements (GHS-US)

:

H220 - EXTREMELY FLAMMABLE GAS

H280 - CONTAINS GAS UNDER PRESSURE; MAY EXPLODE IF HEATED
OSHA-H01 - MAY DISPLACE OXYGEN AND CAUSE RAPID SUFFOCATION
CGA-HG04 - MAY FORM EXPLOSIVE MIXTURES WITH AIR
CGA-HG08 - BURNS WITH INVISIBLE FLAME

Precautionary statements (GHS-US)

:

P202 - Do not handle until all safety precautions have been read and understood
P210 - Keep away from Heat, Open flames, Sparks, Hot surfaces. - No smoking
P271+P403 - Use and store only outdoors or in a well-ventilated place
P377 - Leaking gas fire: Do not extinguish, unless leak can be stopped safely
P381 - Eliminate all ignition sources if safe to do so
CGA-PG05 - Use a back flow preventive device in the piping
CGA-PG10 - Use only with equipment rated for cylinder pressure
CGA-PG12 - Do not open valve until connected to equipment prepared for use
CGA-PG06 - Close valve after each use and when empty
CGA-PG02 - Protect from sunlight when ambient temperature exceeds 52°C (125°F)

METHANE, COMPRESSED

Safety Data Sheet P-4618

This SDS conforms to U.S. Code of Federal Regulations 29 CFR 1910.1200, Hazard Communication.

Date of issue: 01/01/1978 Revision date: 10/17/2016 Supersedes: 03/09/2015

SECTION 1. Product and company identification

1.1 Product identifier

Product form	:	Substance
Name	:	METHANE, COMPRESSED
CAS No	:	74-82-8
Formula	:	CH4

1.2 Relevant identified uses of the substance or mixture and uses advised against

Use of the substance/mixture	:	Industrial use. Use as directed.
------------------------------	---	----------------------------------

1.3 Details of the supplier of the safety data sheet

Praxair, Inc.
10 Riverview Drive
Danbury, CT 06810-6268 - USA
T 1-800-772-9247 (1-800-PRAXAIR) - F 1-716-879-2146
www.praxair.com

1.4 Emergency telephone number

Emergency number	:	Onsite Emergency: 1-800-645-4633
------------------	---	----------------------------------

CHEMTREC, 24hr/day 7days/week
— Within USA: 1-800-424-9300, Outside USA: 001-703-527-3887
(collect calls accepted, Contract 17729)

SECTION 2. Hazard identification

2.1 Classification of the substance or mixture

GHS-US classification

Flam. Gas 1 H220
Compressed gas H280

2.2 Label elements

GHS-US labeling

Hazard pictograms (GHS-US)



GHS02

GHS04

Signal word (GHS-US)

:

DANGER

Hazard statements (GHS-US)

:

H220 - EXTREMELY FLAMMABLE GAS

H280 - CONTAINS GAS UNDER PRESSURE; MAY EXPLODE IF HEATED

OSHA-H01 - MAY DISPLACE OXYGEN AND CAUSE RAPID SUFFOCATION

CGA-HG04 - MAY FORM EXPLOSIVE MIXTURES WITH AIR

Precautionary statements (GHS-US)

:

P202 - Do not handle until all safety precautions have been read and understood

P210 - Keep away from Heat, Open flames, Sparks, Hot surfaces. - No smoking

P271+P403 - Use and store only outdoors or in a well-ventilated place

P377 - Leaking gas fire: Do not extinguish, unless leak can be stopped safely

P381 - Eliminate all ignition sources if safe to do so

CGA-PG05 - Use a back flow preventive device in the piping

CGA-PG10 - Use only with equipment rated for cylinder pressure

CGA-PG06 - Close valve after each use and when empty

CGA-PG11 - Never put cylinders into unventilated areas of passenger vehicles

CGA-PG02 - Protect from sunlight when ambient temperature exceeds 52°C (125°F)



Propane

Safety Data Sheet E-4646

according to the Hazardous Products Regulation (February 11, 2015)

Date of Issue: 10-15-1979

Revised date: 08-03-2016

Supersedes: 10-15-2013

SECTION 1: Identification

1.1. Product Identifier

Product form	:	Substance
Name	:	Propane
CAS No.	:	74-98-6
Formula	:	C3H8
Other means of identification	:	Propane, Liquefied Petroleum Gas, n-propane, dimethylmethane, propyl hydride, refrigerant gas R290
Product group	:	Core Products

1.2. Recommended use and restrictions on use

Recommended uses and restrictions	:	Industrial use Use as directed
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1.3. Supplier

Praxair Canada Inc.
1200 – 1 City Centre Drive
Mississauga - Canada L5B 1M2
T 1-905-803-1600 - F 1-905-803-1682
www.praxair.ca

1.4. Emergency telephone number

Emergency number	:	1-800-363-0042 Ca 1 emergency number 24 hours a day only for spills, leaks, fire, exposure, or accidents involving this product. For routine information, contact your supplier or Praxair sales representative.
------------------	---	--

SECTION 2: Hazard identification

2.1. Classification of the substance or mixture

GHS-CA classification

Flam. Gas 1 H220
Liquefied gas II2B0

2.2. GHS Label elements, including precautionary statements

GHS-CA labelling

Hazard pictograms



GHS02 GHS04

Signal word

:

DANGER

Hazard statements

:

EXTREMELY FLAMMABLE GAS
CONTAINS GAS UNDER PRESSURE; MAY EXPLODE IF HEATED
MAY DISPLACE OXYGEN AND CAUSE RAPID SUFFOCATION
MAY CAUSE FROSTBITE
MAY FORM EXPLOSIVE MIXTURES WITH AIR

Precautionary statements

:

Do not handle until all safety precautions have been read and understood
Keep away from heat, hot surfaces, sparks, open flames and other ignition sources. No smoking
Use and store only outdoors or in a well-ventilated area
Leaking gas fire: Do not extinguish, unless leak can be stopped safely
In case of leakage, eliminate all ignition sources

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SAFETY DATA SHEET**BITUMEN**

000003000832

Version 3.1

Revision Date 2018/06/06

Print Date 2018/06/06

SECTION 1. IDENTIFICATION

Product name	:	BITUMEN
Synonyms	:	Mackay River Bitumen, FireBag Hot Bitumen, Dover Bitumen, Bitumen, Off-Spec Bitumen, Sales Oil, FB HB, FHB, MKB, Natural Asphalt
Product code	:	100236
Manufacturer or supplier's details		SUNCOR ENERGY INC. P.O. Box 2844, 150 - 6th Avenue South-West Calgary Alberta T2P 3E3 Canada
Emergency telephone number	Suncor Energy: +1 403-296-3000; Canutec Transportation: 1-888-226-8832 (toll-free) or 613-996-6666; Poison Control Centre: Consult local telephone directory for emergency number(s).	

Recommended use of the chemical and restrictions on use

Recommended use	:	Raw product for oil refineries to produce fuels and other petroleum based organic products.
Prepared by	:	Product Safety: +1 905-804-4752

SECTION 2. HAZARDS IDENTIFICATION**Emergency Overview**

Appearance	viscous liquid
Colour	black
Odour	Tarry odour. "Rotten egg" if H2S present, but odour is an unreliable warning, since it may deaden the sense of smell.

GHS Classification

Skin irritation	:	Category 2
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GHS label elements

Hazard pictograms



Signal word	:	Warning
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SAFETY DATA SHEET**KEROSENE**

000003000579



Version 2.1

Revision Date 2018/06/07

Print Date 2018/06/07

SECTION 1. IDENTIFICATION

Product name	:	KEROSENE
Synonyms	:	Low Sulphur Kerosene, Kerosine, KEROSENE (TYPE 1-K)
Product code	:	101867, 101866

Manufacturer or supplier's details

SUNCOR ENERGY INC.
 P.O. Box 2844, 150 - 6th Avenue South-West
 Calgary Alberta T2P 3E3
 Canada

Emergency telephone number	Suncor Energy: +1 403-296-3000; Canutec Transportation: 1-888-226-8832 (toll-free) or 613- 996-6666; Poison Control Centre: Consult local telephone directory for emergency number(s).
----------------------------	--

Recommended use of the chemical and restrictions on use

Recommended use	:	Kerosene is a refined petroleum distillate suitable for burning in wick lamps and space heaters designed for kerosene.
Prepared by	:	Product Safety: +1 905-804-4752

SECTION 2. HAZARDS IDENTIFICATION**Emergency Overview**

Appearance	Clear liquid.
Colour	Clear and bright
Odour	Hydrocarbon.

GHS Classification

Flammable liquids	:	Category 3
Skin irritation	:	Category 2
Specific target organ toxicity - single exposure	:	Category 3 (Central nervous system)
Aspiration hazard	:	Category 1

GHS label elements

SAFETY DATA SHEET**NAPHTHA (SWEET)**

000003000638



Version 3.0

Revision Date 2019/01/29

Print Date 2019/01/29

SECTION 1. IDENTIFICATION

Product name	:	NAPHTHA (SWEET)
Synonyms	:	Light Naphtha, Heavy Naphtha, Reformer Feed, Platformer Feed, Hydrodesulfurized Naphtha, Hydrotreated Naphtha, ER57.
Manufacturer or supplier's details		SUNCOR ENERGY INC. P.O. Box 2844, 150 - 6th Avenue South-West Calgary Alberta T2P 3E3 Canada
Emergency telephone number		Suncor Energy: +1 403-296-3000; Canutec Transportation: 1-888-226-8832 (toll-free) or 613-996-6666; Poison Control Centre: Consult local telephone directory for emergency number(s).

Recommended use of the chemical and restrictions on use

Recommended use	:	Light and Heavy naphthas are intermediate refinery products used as feedstocks to platformer units for the production of high octane motor gasoline blending components.
Prepared by	:	Product Safety: +1 905-804-4752

SECTION 2. HAZARDS IDENTIFICATION**Emergency Overview**

Appearance	liquid
Colour	Clear to yellow
Odour	Gasoline like.

GHS Classification

Flammable liquids	:	Category 2
Skin irritation	:	Category 2
Germ cell mutagenicity	:	Category 1B
Carcinogenicity	:	Category 1A
Reproductive toxicity	:	Category 2
Specific target organ toxicity - single exposure	:	Category 3 (Central nervous system)

G.6 REFERENCES

CH E 465 Tutorial on PHA, F&EI and CEI

American Institute of Chemical Engineers (2004), Dow's Fire and Explosion Index Hazard Classification Guide, 7th Edition.

American Institute of Chemical Engineers (2004), Dow's Chemical Exposure Index Guide, 1st Edition.

Praxair Material Safety Data Sheet (SDS) for gases. Retrieved March 18, 2020, from

[http://www.praxair.ca/en-ca/resource-library/sds?s=%22k%22:%22%22,%22p%22:1,%22ps%22:10}](http://www.praxair.ca/en-ca/resource-library/sds?s=%22k%22:%22%22,%22p%22:1,%22ps%22:10)

Suncor Energy, Bitumen Safety Data Sheet. Retrieved March 18, 2020, from
http://www.online.petro-canada.ca/datasheets/en_CA/90000124.pdf

Suncor Energy, Kerosene Safety Data Sheet. Retrieved March 18, 2020, from
http://www.online.petro-canada.ca/datasheets/en_CA/w106.pdf

Suncor Energy, Naphtha Safety Data Sheet. Retrieved March 18, 2020, from
http://www.online.petro-canada.ca/datasheets/en_CA/w344.pdf

SLURRY HYDROCRACKER PROJECT

Appendix H - P&ID and HAZOP

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DATE: April 9th, 2020

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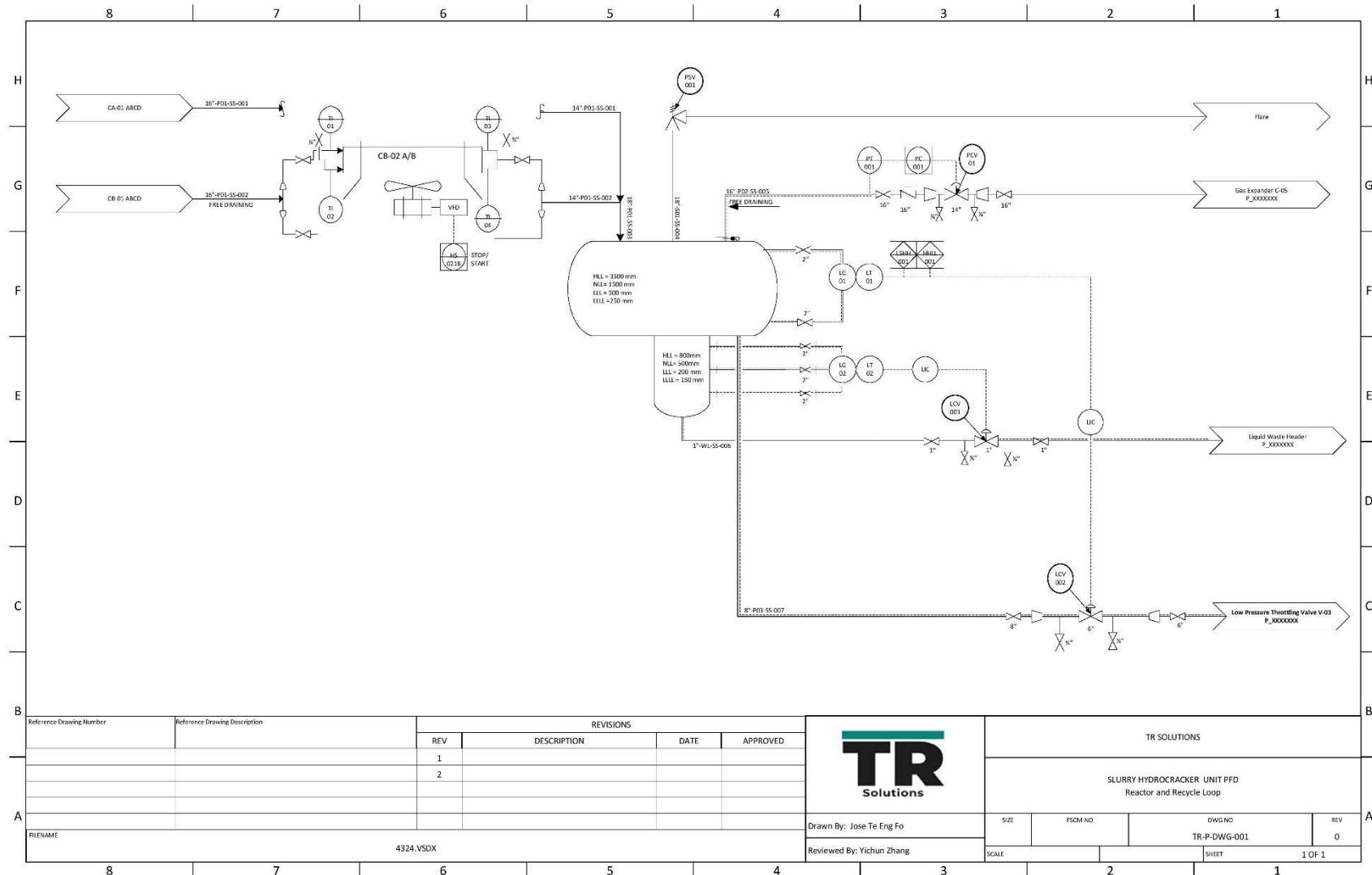
H.1 SUMMARY

This appendix shows a detailed hazard and operability study (HAZOP) on a piping and instrumentation diagram (P&ID). A P&ID for the unit containing the three-phase separator D-03 and the air fin cooler C-01 is shown in Section H2. Section H3 shows the criteria to assess the risks, including a risk matrix and a risk level description table. In Section H3, a HAZOP is performed to examine potential risks associated with deviations in parameters including pressure, temperature, flow, and level.

H.2 PIPING AND INSTRUMENTATION DIAGRAM (P&ID)

Figure H2 shows the P&ID around the three-phase separator D-03 and the air fin cooler C-02. The unit is divided into two nodes. Node 1 is the cooler unit including the stream coming from cooler C-01 and the stream entering the vessel. Node 2 includes the separator vessel, its three outlet streams, and the PSV.

Figure H2. A P&ID of the three-phase separator D-03 and waste heat boiler C-01.



H.3 RISK ASSESSMENT CRITERIA

The HAZOP study of the P&ID is based on the criteria indicated in Table H3.

Table H3. Risk matrix.

H.4 HAZARD AND OPERABILITY (HAZOP) STUDY

HAZOP Study Title: Slurry Bitumen Hydrocracker

P&ID Description: Three-Phase Separator D-03

Node #: 1

Node Description: Air fin cooler

Item #	Guide -word	Parameter	Deviation	Possible Causes	Consequences	CAT	Risk without Existing Safeguards			Safeguards	Risk with Existing Safeguards			Recommendations / Actions
							S	L	RR		S	L	RR	
1	High	Temperature	Feed enters the unit at increased temperature.	Upstream WHB fails to cool down the stream to designed temperature.	Separator operates at higher temperature. Undesired separation may be achieved.	Assets	3	4	6	TIC-002 combined with TT-02 monitors the temperature of the fluid exiting the air fin cooler. If the temperature is higher than designed, the VFD will increase the air flow rate to provide a higher cooling rate.	1	2	2	
2	Low	Temperature	Feed enters the unit at decreased temperature.	Upstream WHB cooled down the stream to a temperature lower than designed.	Separator operates at lower temperature. Undesired separation may be achieved.	Assets	3	4	6	TIC-002 combined with TT-02 monitors the temperature of the fluid exiting the air fin cooler. If the temperature is lower than designed, the VFD will reduce the air flow rate to provide a lower cooling rate.	1	2	2	

Node #: 2

Item #	Guide-word	Parameter	Deviation	Possible Causes	Consequences	CAT	Risk without Existing Safeguards			Safeguards			Risk with Existing Safeguards			Recommendations / Actions
							S	L	RR	S	L	RR	S	L	RR	
1	High	Level	Liquid hold-up height in the liquid product separator is increased.	Upstream process sends a higher flow rate of fluid into the vessel.	Liquid-gas engagement volume is decreased, potentially leading to insufficient separation. Liquid droplets can enter turbine EX-01 and cause damages to the equipment.	Assets	4	3	6	LG-02 of the vessel will monitor the liquid level. If it exceeds the HLL, LCV-01 at the bottom of the vessel will be opened to increase liquid draw out of the vessel until liquid level returns to the normal range.			1	2	2	
2	Low	Level	Liquid hold-up height in the liquid product separator is decreased.	Upstream process sends a lower flow rate of fluid into the vessel.	Loss of liquid hold-up in the vessel. Unseparated fluid may bypass at the bottom. Product may not meet specifications.	Assets	4	3	6	LG-02 of the vessel will monitor the liquid level. If it drops below the LLLL, and LCV-01 at the bottom of the vessel will be closed to reduce liquid draw out of the vessel until liquid level returns to the normal range.			1	2	2	
3	High	Pressure	Pressure is built-up in the vessel.	Downstream piping valve is blocked.	Vessel burst. Flammable and high-pressure fluid is released, potentially leading to fire or explosion hazard.	Workers and the public	6	3	8	1. PSV at the top of the vessel will be opened to release content of the vessel to flare until the pressure is returned to the normal range. 2. Emergency shutdown.			4	2	5	1. Workers on site must wear PPE all the time. 2. Provide adequate training and accessible procedure for emergency shutdown. 3. Prepare an evacuation plan.
4	High	Pressure	Vessel operates at an elevated pressure.	Pressure fluctuations due to changes in the feed and liquid draw flow rates.	The pressure of the fluid being sent to downstream expander EX-01 exceeds normal operating range. It may lead to undesirable expansion outcomes.	Assets	4	3	6	PT-01 will monitor the pressure of the gas and PCV-01 will maintain a normal range of pressure.			2	2	3	
5	High	Pressure	Vessel operates at an elevated pressure.	Pressure fluctuations due to changes in the feed and liquid draw flow rates.	The pressure of the fluid being sent to downstream expander EX-01 exceeds the normal operating range. It may lead to undesirable expansion outcomes.	Assets	4	3	6	PT-01 will monitor the pressure of the gas and PCV-01 will maintain a normal range of pressure.			2	2	3	

6	Low	Pressure	Vessel operates at a reduced pressure.	Pressure fluctuations due to changes in the feed and liquid draw flow rates.	The pressure of the fluid being sent to downstream expander EX-01 is below the normal operating range. It may lead to undesirable expansion outcomes.	Assets	4	3	6	PT-01 will monitor the pressure of the gas and PCV-01 will maintain a normal range of pressure.	2	2	3	
7	High	Level	Liquid hold-up height in the water separator is increased.	Upstream process sends a higher flow rate of fluid into the vessel.	Liquid-gas engagement volume is decreased, potentially leading to insufficient separation. Water content in the liquid product may increase and product value may be downgraded.	Assets	3	3	5	LG-02 of the water separator will monitor the liquid level. If it exceeds the HLL, LCV-002 will be opened up to increase liquid draw out of the vessel until liquid level returns to the normal range.	1	2	2	
8	Low	Level	Liquid hold-up height in the water separator is decreased.	Upstream process sends a lower flow rate of fluid into the vessel.	Loss of liquid hold-up in the vessel. Unseparated fluid may bypass at the bottom, leading to loss of production.	Assets	3	3	5	LG-02 of the water separator will monitor the liquid level. If it drops below the LLLL, LCV-002 will be closed to reduce liquid draw out of the vessel until liquid level returns to the normal range.	1	2	2	

Node Description: Three-phase separator

H.5 REFERENCES

CHE 464 and 465 Class Notes

SLURRY HYDROCRACKER PROJECT

Appendix I – Economic Analysis

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DATE: April 9th, 2020

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I.1 SUMMARY

This appendix contains all the data, calculation and assumptions for the economic analysis. The capital costs sample calculation contains all the equipment cost calculations as well as class five estimations on capital cost. The operating costs sample calculation contains sample calculations for utility cost calculations, diluent cost calculations, operating labour cost calculations, waste disposal cost calculations. Finally, in the discounted cash flow sample calculation section, a cash flow for the first operating year (2025) is illustrated, a net present value (NPV) is and sensitivity analysis is obtained.

All prices related to the operating cost is taken from the 2019 average price as the price for January 2020. A currency rate of 1.34 CAD = 1 USD is used. To account for the location from the U.S coast in U&V (2004) to Edmonton, a location factor of 1.35 is applied. CE Plant Cost Index value 620 for 2020 is used. An operating factor of 0.9 is applied.

I.2 CAPITAL COSTS

Table 1 shows the capital cost estimation including the bare module costs for equipment cost and class five estimations. This table also includes estimations for construction DFL adjustments, indirect project expenses, auxiliary facilities etc. These estimations are done by estimating the cost to be a percentage of the total capital cost. The total grass-root capital cost is determined to be 6.9 billion dollars.

Table I1. (the capital cost table from capital cost spread sheet)

Equipment category	Equipment Number	Capacity or size specifications	Purchased Equipment Cost CP	Material factor FM	Pressure or other factors FP	Actual bare module factor FaBM	Actual bare module cost CaBM (2004)	Cost Index Adjustment CBM (2020)	Currency Adjustment CBM(\$CAD)	Location factor	Number of Trains	Number of equipment / train	Total
Reactors	R-A01 - R-B08	vertical D=4m L=30.8m	\$160,000.00	2.5	7.2	33.4	\$ 5,344,000.00	\$ 8,283,200.00	\$ 11,099,488.00	\$ 14,984,308.80	2	4	\$119,874,470.40
	R-05 & R-06	Vertical D=4m L=19.7m	\$116,000.00	2.5	3.9	9.1	\$ 1,049,878.00	\$ 1,627,310.90	\$ 2,180,596.61	\$ 2,943,805.42	1	2	\$5,887,610.84
Stripper Column	T-02	D=1.5m H=4.04m	\$14,000.00	2.5	2.1	13	\$ 182,000.00	\$ 282,100.00	\$ 378,014.00	\$ 510,318.90	1	1	\$510,318.90
Two-Phase Separators	D-01	vertical D = 4m L = 20m	\$117,445.00	2.5	7.46	34.60	\$ 4,064,148.99	\$ 6,299,430.94	\$ 8,441,237.46	\$ 11,395,670.56	2	2	\$45,582,682.26
	D-02	vertical D = 4m L = 20m	\$117,445.00	2.5	7.43	34.46	\$ 4,047,647.97	\$ 6,273,854.35	\$ 8,406,964.83	\$ 11,349,402.52	2	2	\$45,397,610.09
	D-04	vertical D = 1.5m L = 7.5m	\$23,972.35	2.5	4.61	22.21	\$ 532,473.84	\$ 825,334.45	\$ 1,105,948.16	\$ 1,493,030.02	1	1	\$1,493,030.02
	D-05	vertical D = 4m L = 20m	\$117,445.00	2.5	2.18	11.91	\$ 1,398,922.63	\$ 2,168,330.07	\$ 2,905,562.30	\$ 3,922,509.10	1	4	\$15,690,036.42
	D-06	vertical D = 4m L = 20m	\$117,445.00	2.5	0.91	6.44	\$ 756,087.42	\$ 1,171,935.50	\$ 1,570,393.57	\$ 2,120,031.32	1	2	\$4,240,062.65
Three-Phase Separators	D-03	horizontal D=3m L=7.08m	\$26,000.00	2.5	7.39	28.81	\$ 749,161.40	\$ 1,161,200.17	\$ 1,556,008.23	\$ 2,100,611.11	1	1	\$2,100,611.11
	D-07	horizontal D=1m L=1.1m	\$3,500.00	2.5	4.64	18.69	\$ 65,408.35	\$ 101,382.94	\$ 135,853.14	\$ 183,401.74	1	1	\$183,401.74
Compressors	CT-C01	centrifugal 26777 kW	\$14,754,029.98	2.5			\$ 36,885,074.95	\$ 57,171,866.17	\$ 76,610,300.67	\$ 103,423,905.91	1	1	\$103,423,905.91
	CT-C02	centrifugal 27957 kW	\$15,351,488.28	2.5			\$ 38,378,720.70	\$ 59,487,017.09	\$ 79,712,602.89	\$ 107,612,013.91	1	1	\$107,612,013.91
	CP-01	centrifugal 480 kW	\$363,487.53	2.5			\$ 908,718.83	\$ 1,408,514.18	\$ 1,887,409.00	\$ 2,548,002.15	1	1	\$2,548,002.15
	CP-02	centrifugal 9581 kW	\$5,728,918.00	2.5			\$ 14,322,295.00	\$ 22,199,557.25	\$ 29,747,406.72	\$ 40,158,999.07	1	1	\$40,158,999.07
Feed Pumps	P-01A	centrifugal capacity = 0.104 m3/s ws = 769 kW ps = 5 bar pd = 61 bar	\$7,800.00	1.4	1	4	\$ 31,200.00	\$ 48,360.00	\$ 64,802.40	\$ 87,483.24	2	1	\$174,966.48
	P-01B	centrifugal capacity = 0.104 m3/s ws = 769 kW ps = 61 bar pd = 116 bar	\$7,800.00	1.4	2.3	6.6	\$ 51,480.00	\$ 79,794.00	\$ 106,923.96	\$ 144,347.35	2	1	\$288,694.69

	P-01C	centrifugal capacity = 0.104 m ³ /s ws = 769 kW ps = 116 bar pd = 172 bar	\$7,800.00	1.4	2.6	7.3	\$ 56,940.00	\$ 88,257.00	\$ 118,264.38	\$ 159,656.91	2	1	\$319,313.83
	P-02	centrifugal capacity = 0.0016 m ³ /s ws = 0.847 kW ps = 0.9 bar pd = 4.9 bar	\$3,500.00	1.4	1	4	\$ 14,000.00	\$ 21,700.00	\$ 29,078.00	\$ 39,255.30	1	1	\$39,255.30
Coolers	C-01	A=653.4 m ²	\$61,184.37	1	1.2	3.3	\$ 201,908.42	\$ 312,958.05	\$ 419,363.79	\$ 566,141.12	2	1	\$1,132,282.23
	CT-C01	A = 161.4 m ²	\$34,942.36	1	1.08	3.2	\$ 111,815.55	\$ 173,314.11	\$ 232,240.90	\$ 313,525.22	1	1	\$313,525.22
Fire heaters	F-01	Q = 31627.5 kW	1.20E+06		1.35	2.1	\$ 2,520,000.00	\$ 3,906,000.00	\$ 5,234,040.00	\$ 7,065,954.00	2	1	\$14,131,908.00
	F-02	Q = 56430.6 kW	2.00E+06		1.35	2.1	\$ 4,200,000.00	\$ 6,510,000.00	\$ 8,723,400.00	\$ 11,776,590.00	2	1	\$23,553,180.00
Exchanger	E-01	A=1790, Floating Head	45000	1.7	1.3	4.8	\$ 216,000.00	\$ 334,800.00	\$ 448,632.00	\$ 605,653.20	2	1	\$1,211,306.40
	E-02	A=59.0880257 U tube	6000	1.7	1.3	4.8	\$ 28,800.00	\$ 44,640.00	\$ 59,817.60	\$ 80,753.76	2	1	\$161,507.52
Waste Heat Boiler	C-01	A=7396.1 m ²	55000	1.7	1.15	4.5	\$ 247,500.00	\$ 383,625.00	\$ 514,057.50	\$ 693,977.63	2	4	\$5,551,821.00
Airfin Coolers	C-02	A=3653.278 m ² 1/train	90000	3	1.15	6.5	\$ 585,000.00	\$ 906,750.00	\$ 1,215,045.00	\$ 1,640,310.75	2	1	\$3,280,621.50
	C-03	A=83.62918 m ²	40000	3	1.15	6.5	\$ 260,000.00	\$ 403,000.00	\$ 540,020.00	\$ 729,027.00	1	1	\$729,027.00
	C-04	A=1637.521 m ²	80000	3	1.15	6.5	\$ 520,000.00	\$ 806,000.00	\$ 1,080,040.00	\$ 1,458,054.00	1	1	\$1,458,054.00
	C-04	A=1794.176 m ²	82000	1	1	4	\$ 328,000.00	\$ 508,400.00	\$ 681,256.00	\$ 919,695.60	1	1	\$919,695.60
Gas Turbine	EX-01	Ws = 2640 kW	\$328,901.76			3.5	\$ 1,151,156.16	\$ 1,784,292.05	\$ 2,390,951.34	\$ 3,227,784.31	1	1	\$3,227,784.31
Amine Plant	-	Capacitcy= 293.3 MMSCFD	-	-	-	-	-	-	-	-	-	-	\$91,911,973.09
SMR	-	Capacitcy= 1.73E+7 m ³ /d	-	-	-	-	-	-	-	-	-	-	\$589,214,198.17
Sulfur Plant SRU	-	Capacitcy= 1514 LT/d	-	-	-	-	-	-	-	-	-	-	\$82,021,055.27
DRU	-	Capacity=130,000BPS D	-	-	-	-	-	-	-	-	-	-	\$147,456,925.69
Zinc Oxide Bed	-	Capacity=8443.2 m ³ /h	-	-	-	-	-	-	-	-	-	-	\$4,114,854.65
Total Cp (with CI & currency adjustment):													\$1,465,914,705.40

Table I2. Facility cost additions.

Total Cp (with CI & currency adjustment)		\$1,465,914,705.4 0
Operating Facility	33%	\$483,751,852.78
% DFL in Turnaround	58%	\$842,900,955.61
% DFL in Winter	30%	\$439,774,411.62
Total DFL		\$1,766,427,220.0 1
Overhead for DFL additions	0.7*C L	\$728,559,608.58
Misc - Total CAPEX in \$M	10%	\$176,642,722.00
CBM Facility Bare Module Capital		\$2,671,629,550.5 9
Process Contingency	40%	\$1,068,651,820
Contractual Fee	3%	\$80,148,887
CTM Facility Total Module Capital		\$3,820,430,257
Site development	5%	\$191,021,513
Auxiliary Buildings	4%	\$152,817,210
Off-Site facilities	21%	\$802,290,354
Total Auxilliary Facilities	30%	\$1,146,129,077
Sub Total Grass Roots Facility		\$4,966,559,335
Project Contingency	20%	\$993,311,867
Sub Total Grass Roots Facility and Contingency		\$5,959,871,201
Engineering / Procurement	10%	\$536,388,408
Owners' Cost	3%	\$178,796,136
Commissioning & Startup	4%	\$238,394,848
CGR Total Grass-roots Capital (Total Fixed Capital)		\$6,913,450,594

A summary of costs regarding the equipment cost and class five capital cost estimation can be found in Table 2 and Table 3.

Table I3. Equipment cost summary.

Equipment Name	Number of Equipment	Total Cost
Reactors	8	\$125,762,081
Separators	17	\$114,687,434
Pumps	3	\$822,230
Compressors	4	\$253,742,921.0 3
Coolers	3	\$1,445,807
Fire Heaters	4	\$37,685,088
Heat Exchangers	4	\$1,372,814
Waste Heat Boilers	8	\$5,551,821
Expander	1	\$3,227,784
Air Fin Coolers	5	\$6,387,398
Total:		\$551 M

Table I4. Class five capital cost summary.

Equipment Name	Cost
Hydrocracker Unit	\$551 M
Amine Treatment Unit	\$92 M
Hydrogen Plant (SMR)	\$589 M
Sulfur Plant	\$82 M
Diluent Recovery Unit	\$147 M
Zinc Oxide Bed	\$4 M

I.3 OPERATING COSTS

Table 4 contains all the steps in reaching the total annual expenses of this project. This includes utility costs, raw materials cost, catalyst cost, operating costs etc. Indirect costs including overhead, tax and insurance etc are also estimated to be a percentage of total capital cost.

Table I5. Manufacturing cost summary.

Capital		
	\$/yr	\$/yr
Manufacturing expenses		
Direct		
Raw materials (Diluted Bitumen + Natural Gas + Process Steam)	\$ 2,364,319,537	
Catalysts and solvents (Iron Catalyst @ 6655.16 kg/h)	\$ 262,346,407	
Operating labour	\$ 1,507,322	
Supervisory and clerical labour (10% of operating labour)	\$ 150,732	
Utilities		
Electricity, 597330662.3 kWh @ \$0.127/kWh	\$ 80,507,322	
Natural Gas, 17862588688GJ @ \$1.4/GJ	\$ 22,499,449	
Demineralized Water 597330662.3 kWh @ \$5.46/kWh	\$ 8,400,098	
Process Steam 35478000 kg @ \$0.022/m ³	\$ 26,176,751	
Cooling Water 2910884.844 m ³ @\$0.0507 /m ³	\$ 193,610	
Zinc Oxide Bed (Class V)	\$31,611,306.00	
Maintenance and repairs (2% of fixed capital)	\$ 138,269,012	
Operating supplies (10% of maint. & repairs)	\$ 13,826,901	
Laboratory charges (10% of operating labour)	\$ 57,821,148	
Total, ADME	\$ 3,323,494,670	\$ 3,323,494,670
Indirect Costs		
Overhead (payroll and plant), packaging, storage (50% op. labour, supervision and maintenance)	\$ 289,181,107	
Local taxes (2% of fixed capital)	\$ 138,269,012	
Insurance (1% of fixed capital)	\$ 69,134,506	
Total, AIME	\$ 496,584,625	\$ 496,584,625
Total manufacturing expense, AME = ADME + AIME		\$ 3,820,079,295
General expenses		
Administrative costs (25% of overhead)	\$ 72,295,277	
Total general expense, AGE	\$ 72,295,277	\$ 72,295,277
Depreciation (approx. 10% of fixed capital), ABD		\$ 691,345,059
Total expenses , ATE		\$ 4,583,719,631
Revenue from sales (kg/yr @ \$/kg), AS		\$ 4,018,036,780
Net annual profit, ANP		\$ (565,682,851)
Income taxes (net annual profit tax rate), AIT		\$ (152,734,370)
Net annual profit after taxes (ANP-AIT), ANNP		\$ (412,948,481)

Table I6. Utility cost summary.

Utility Type	Consumption	Unit	Unit Price (CAD)	Annual Cost (CAD)
Electricity	659829380	kwh	0.12201233	68,277,548
Natural Gas (Utility)	16061474.65	GJ	1.40083333	6,978,847
Demineralized Water	1,538,304	m3	5.46062341	8,400,098
Process Steam	1283520039	kg	0.02039450	26,176,750
Cooling Water	2910884.844	m3	0.06651250	193,610

Table I7. Summary of operating consumption.

Name	Price (CAD)
Bitumen	\$1.64
Natural Gas	\$0.02
Catalyst	\$0.26
Other Direct Costs	\$0.17
Utilities	\$0.11
Indirect	\$0.33
General Expenses	\$0.03
Bitumen	\$1.64
Naphtha	\$0.7
Raw Materials	\$0.03

I.4 FEED COSTS

The battery-limit feed cost to the process consists of three parts; bitumen, diluent and natural gas. Bitumen is the raw material that needs to be upgraded to crude oil. It is usually transported by pipe and mixed with diluent, with a ratio of 30 vol% of diluent and 70 vol% of bitumen. The natural gas is consumed in the SMR unit for hydrogen production.

The diluent is naphtha, but when it is used as diluent, the price is lower than naphtha as a chemical selling in the market. Refer to page I21, the price for diluent is calculated in order to prevent using the price for naphtha and to save some cost.

Table I8. Battery limit feed cost summary.

Feed	Prices (CAD)
Diluent	\$ 699,100,792
Bitumen	\$ 1,641,030,120
Natural Gas	\$ 24,188,624.46

I.5 REVENUE

The revenue consists of three parts, the diluent recovered from the DRU is assumed to be pure naphtha available for sale. Sulphur recovered from the Claus plant is marketed as a product. The crude oil is priced as the WTI Cushing price. All prices are the average price in 2019.

Table I9. Summary of Annual Revenue

Name	Unit Price	Annual Revenue (CAD)
Crude Oil	475.83 CAD/m ³	\$ 3,068,692,528.33
Naphtha	427.66 CAD/m ³	923804667.03
Sulphur	167.5 CAD/m ³	\$ 25,539,584.59

I.6 CLASS V OPERATING COST ESTIMATES

The class five operating cost estimations is done through research, utilities for each of class five units is obtained and added to the total annual utility cost in the operating cost table.

The class five operating estimation for the zinc oxide bed is done by adding the cost for zinc oxide catalyst and the solid waste disposal together to get the annual expense. According to Hassan, Khammas, and Al-Mayah (2008), the price for zinc oxide is consist of 90 wt% of zinc oxide and 10 wt% of alumina, and the efficiency of the hydrogen sulphide removal reaction is assumed to be 90%. Due to lack of information, the price of zinc oxide and alumina is obtained from an online trading website called alibaba. The suppliers of zinc oxide and alumina from this market mostly ship their product f.o.b. with a range of prices listed, for example, for most of the

suppliers, \$2000 - 3000 USD per ton of zinc oxide powder is the normal price. The lower bound, \$2000 USD/t is chosen since this project has a big consumption on zinc oxide catalyst, it is reasonable to assume that a discounted price can be given to the project with some negotiation.

I.7 CASH FLOW ANALYSIS

The evaluation of this project involves a 40 years of cash flow. The first five years will be the construction of the grass-root plant site. With a total fixed capital of \$ 6,898,227,929 dollars to be installed each year. The working capital is estimated as 15% of total fixed capital invested at the beginning of the 5th year. A schedule of investment instalment is listed in the following table:

Table I10. Investment instalments in the construction period.

Year	Fixed Capital	Percent of Total Fixed Capital
0	\$ (344,911,396.47)	5%
1	\$ (1,034,734,189.42)	15%
2	\$ (1,379,645,585.89)	20%
3	\$ (1,379,645,585.89)	20%
4	\$ (2,759,291,171.78)	40%

The reason why most of the instalments are installed to the last three years is (Appendix K) because for most of the projects, the first two years are mainly works on project scope, detail engineering etc. In the last three years of development the project is going to consume most of the investment including paying for bulk of equipment, mechanical constructing, building the steel wall equipment, building the piping systems and finishing the work etc.

The net present value by the end of the 35th operating year was determined to be \$(-5,945,636,564) dollars.

The return on capital of this project is 0.11%, which is smaller than the hurdle rate of 15%. Which means this project will not be executed from an economic point of perspective.

I.7.1 Sensitivity Analysis

A sensitivity analysis is done on main factors that will significantly affect the cash flow and NPV of the project. These factors are bitumen price, crude oil price, total capital and discount factor. The values of these factors had been changed as an input to get the resultant NPV. The way these values had been adjusted is illustrated in the following table, and the sensitivity analysis of each factor is illustrated in the next figure.

Table I11. Factor adjustments for sensitivity analysis.

	Bitumen	WTI Cushing	Capital	Discount Rate
Increase	20%	20%	50%	at 20%
Decrease	-20%	-20%	-20%	at 10%

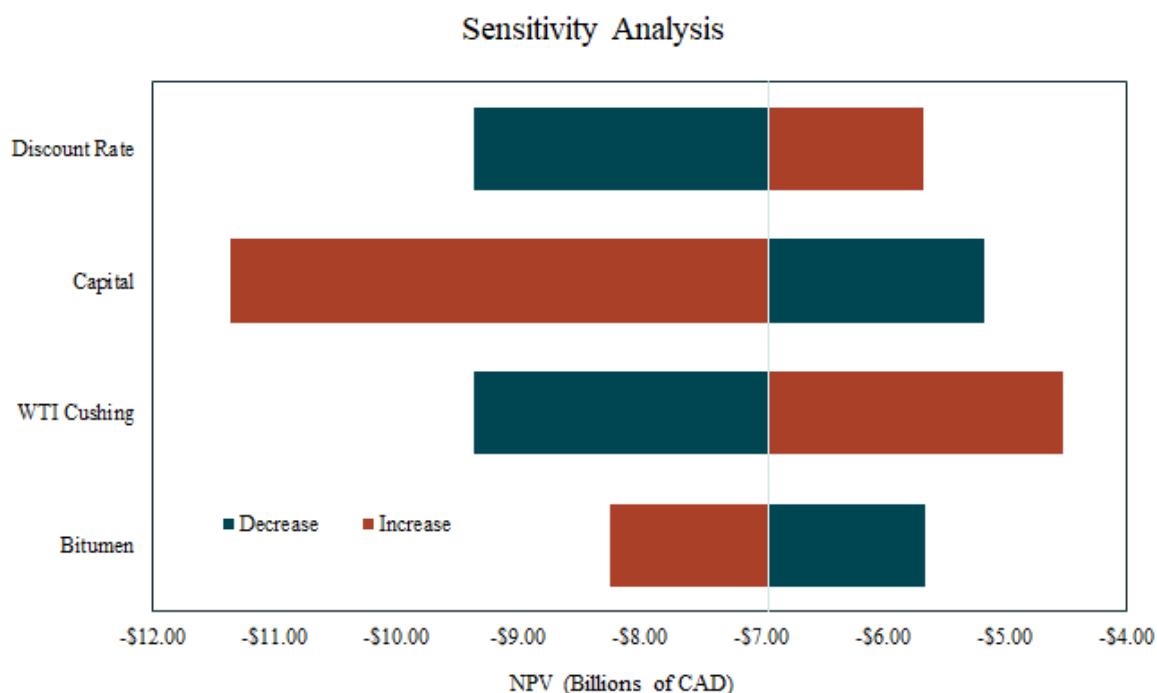


Figure I11. Sensitivity analysis.

I.8 SAMPLE CALCULATIONS

The following is sample calculations by hand for calculations related to capital cost, operating cost, and cash flow. The purpose for this section is to have a detailed demonstration on how the calculations in the economic excel sheet is done. The results from these hand calculations might be slightly different compared to the value in the excel sheet, due to calculation error, but the values should be consistent in general.

I.8.1 Capital Cost Sample Calculation



Slurry Reactor R-01 Cost using U&V

Basis:

- Vertical vessel
- Location factor Edmonton 1.35
- Design P = 250 bar
- CEI = 620 (2020)
- Design T = 467 °C
- CEI = 400 (2004)
- Diameter = 4m
- Currency conversion 1.34
- Length (tangent to tangent) = 30.8 m
- Material of construction: Carbon Steel clad with stainless steel

Using U&V Figure 5.44 pg. 387 - 388

$$C_p = \$160,000$$

$$\bar{F}_M = 2.5$$

$$\bar{F}_P = 7.2$$

$$\bar{F}_P \times \bar{F}_M = 2.5 \times 7.2 = 18$$

Bare module factor, Fig 5.46 $\bar{F}_{BM}^a = 33.4$

$$C_{BM} = C_p \times \bar{F}_{BM}^a = \$160,000 \times 33.4$$

$$C_{BM} = \$5.3 \text{ M USD, USGC, 2004}$$

$$C_{BM} = \$5.3 \text{ M} \times \frac{620}{400} \times 1.35 \times 1.34$$

$$C_{BM, 2020} = \$14.9 \text{ M CAD}$$

The rest of the F reactors are costed in the same manner.



Two-Phase Separator Costing

ex) Flash Drum D-01 (vertical)

According to Ulrich & Vasudevan (2004), Figures 5.44-5.45

For $L = 20\text{m}$, $D = 4\text{m}$, $C_p = \$1.17 \times 10^5$.

Stainless clad was chosen to avoid corrosion

$$F_M = 2.5$$

$$P = 168 \text{ bar} \rightarrow F_P = 7.5$$

$$F_P \times F_M = (2.5)(7.5) = 18.75 \rightarrow F_{BM}^a = 35$$

$$C_{BM}^a = (F_{BM}^a)(C_p) = (35)(\$1.17 \times 10^5) = \$4.06 \times 10^6$$

The rest of the two phase separators are costed in the same manner.

Pump Costing

ex) Pump P-02 (centrifugal)

According to Ulrich & Vasudevan (2004), Figures 5.49-5.51

At $W_s = 0.8467 \text{ kW}$, $C_p = \$3500$

At $P_s = 90 \text{ kPa}$, $F_P = 1.0$

Cast steel was chosen to avoid corrosion. $\rightarrow F_M = 1.4$

$$F_P \times F_M = 1.4 \rightarrow F_{BM}^a = 4$$

$$C_{BM}^a = (F_{BM}^a)(C_p) = (4)(\$3500) = \$14,000$$

The rest of the pumps are costed in the same manner.



Compressor Costing

ex) Compressor CT-CP01 (centrifugal).

According to VMG Symmetry, the compressor duty required to achieve the pressure difference between inlet & outlet is $W_s = 26777 \text{ kW}$.

The equipment type was decided to be centrifugal.

Centrifugal compressors typically have an efficiency $\epsilon_i = 50 \sim 85\%$ (Ulrich & Vasudevan, 2004)

Assuming $\epsilon_i = 0.75$,

$$W_f = \epsilon_i W_s = (0.75)(26777 \text{ kW}) = 20083 \text{ kW}$$

According to Ulrich & Vasudevan, (2004), Figure 5.30,

$$C_p = \$1.48 \times 10^7$$

The risk of corrosion is low, so carbon steel material is selected.

$$F_{BM} = 2.5 \quad (\text{Ulrich \& Vasudevan, 2004})$$

$$C^a_{BM} = (F_{BM})(C_p) \quad (\text{Ulrich \& Vasudevan, 2004})$$

$$= (2.5)(\$1.48 \times 10^7)$$

$$= \$3.69 \times 10^7$$

The rest of the compressors are costed in the same manner.



Heater Costing

ex) Fired Heater F-01

According to VMG Symmetry, the duty required to achieve the desired temperature is $\dot{Q} = 31627.5 \text{ kW}$

According to Ulrich & Vasudevan (2004), Figure 5.27

At $\dot{Q} = 31627.5 \text{ kW}$, C_p of a non-reactive heater is $\$1.20 \times 10^6$.

At $P = 171 \text{ bar}$, $F_p = 1.35$

Carbon steel was chosen because the risk of corrosion is low. $F_{BM} = 2.1$

$$C^a_{BM} = (F_{BM}) (C_p) = (2.1) (\$1.20 \times 10^6) = \$2.52 \times 10^6$$

The rest of the heaters are costed in the same manner.



3 phase separator costing

a) Horizontal 3-phase separator D-03

According to Ulrich & Vasudevan (2004) Figure

For $D = 3\text{ m}$, $L = 7.08\text{ m}$

$$C_p = 26,000 \text{ USD}$$

Carbon Steel was chosen $\rightarrow F_M = 2.5$

$$P = 16700 \text{ kPa} \rightarrow F_p = 7.8$$

$$F_{BM}^a = F_p \cdot F_M = 7.8 \times 1.67 = 13$$

$$C_{BM}^a = C_p \cdot F_{BM}^a = 26,000 \times 13 = 338,000 \text{ USD}$$

Inflation & currency

$$C_{BM}(2020) = C_{BM}^a \times \frac{620}{400} \times \frac{1.34 \text{ CAD}}{1 \text{ USD}} \times 1.35 \text{ (locational factor)}$$

$$= 947,735 \text{ CAD in 2020.}$$

Another 3 phase separator is costed in the same manner.



Class 5 → Capital cost estimation

M = million

All the class 5 plants cost estimation used the Capacity ratio estimates (6/10th rule)

CEI (2020) = 620

Location - 1.35

Currency - 1.34

$$\frac{\text{Cost A}}{\text{Cost B}} = \left(\frac{\text{Capacity A}}{\text{Capacity B}} \right)^n$$

n ≈ depends on the type of process plant scale exponent

Amine treater plant

Cost B = \$ 15 M, USD, 1991 (Maples)

Capacity B = 100 MMSCFD

CEI (1991) = 360

Stream factor = 0.95

n = 0.6

Capacity A = 293.3 MMSCFD

$$\text{Cost A} = \$ 15 \text{M} \times \left(\frac{293.3}{100 \times 0.95} \right)^{0.6} \times \frac{620}{360} \times 1.35 \times 1.34$$

Cost A = \$ 91.9 M (CAD, January 2020)

Sulfur Plant (SRU)

Cost B = \$ 5 M, USD, 1991 (Maples)

Capacity B = 100 LT/D stream factor = 0.95

Capacity A = 1513.92 LT/D n = 0.6

$$\text{Cost A} = \$ 5 \text{M} \times \left(\frac{1513.92}{100 \times 0.95} \right)^{0.6} \times \frac{620}{360} \times 1.34 \times 1.35$$

Cost A = \$ 82 M (CAD, January 2020)



Diluent recovery unit (DRU)

Cost B = \$ 38 M, USD, 1991 (Maples)

Capacity B = 100,000 BPSD stream factor = 0.95

Capacity A = 130,000 BPSD n = 0.7

$$\text{Cost A} = \$ 38 \text{ M} \times \left(\frac{130,000}{100,000 \times 0.95} \right)^{0.7} \times \frac{620}{360} \times 1.34 \times 1.35$$

Cost A = \$ 147 M (Cad, January 2020)

Steam Methane Reformer (SMR)

Cost B = \$ 50 M, USD, 1995 (Murry Gray)

Capacity B = $1.7 \times 10^6 \text{ m}^3/\text{day}$ CEI (1995) = 382.9

Capacity A = $1.73 \times 10^7 \text{ m}^3/\text{day}$ n = 0.6

$$\text{Cost A} = \$ 50 \text{ M} \times \left(\frac{1.73 \times 10^7}{1.7 \times 10^6} \right)^{0.6} \times \frac{620}{382.9} \times 1.34 \times 1.35$$

Cost A = \$ 589 M (Cad, January 2020)

Zinc Oxide Bed

The zinc oxide bed is estimated to be 2 vertical vessels in parallel

The sizing calculation is shown in Appendix E excel sizing sheet "vessels"

Capacity = $8443.2 \text{ m}^3/\text{h}$

L = 17.9 m Material → Carbon steel

D = 4 m

P = 26 bar



Using U&V Fig 5.44, 5.45, 5.46

$$C_p = \$108,000$$

$$F_M = 1$$

$$F_p \times F_M = 2.5$$

$$F_p = 2.5$$

Barn module factor $F_{BM}^a = 6.9$

$$C_{BM} = C_p \times F_{BM}^a = \$108,000 \times 6.9 = \$745200$$

C_{BM} = \$745200 USD, USGC, 2004

$$C_{BM\ 2020} = \$745200 \times \frac{600}{400} \times 1.35 \times 1.34$$

$$C_{BM\ 2020} = \$2M (\text{Cad}, 2020)$$

For the 2 vessels

$$C_{BM} = \$2M \times 2 = \$4M (\text{CAD}, 2020)$$



Expander Costing

Equipment: Gas Turbine EX-01

According to sizing calculations (sample calculations shown in Appendix E)

$$W_s = 2640 \text{ kW}$$

According to Ulrich & Vasudevan (2004, Figure 5.21)

At $W_s = 2640 \text{ kW}$, C_p of a gas turbine = $\$3.3 \times 10^5$

FBM of a gas turbine = 3.5

$$C_{BM}^a = (3.5)(3.3 \times 10^5) = \$1.15 \times 10^6$$

I.8.2 Utilities Unit Cost Sample Calculation



Utility unit cost sample calculation

According to Ulrich & Visudhevam (2004) P416-418

- price per unit of Utility

$$C_{\text{unit}} = a \cdot CE \text{ plant cost index} + b \cdot \text{Fuel price}$$

- where a, b are constants given in the book,
 a, b depend on flow rates etc.

- CE plant cost index in 2020 = 620

- Natural gas is selected as fuel and
 according to the government of Alberta the price
 in 2020 is $1.4 / 1.34 = 1.045 \text{ USD/GJ}$

- Electricity

$$a = 0.00013 \quad b = 0.01$$

$$C_{\text{unit}} = 0.00013 \times 620 + 1.045 \times 0.01 = 0.091 \text{ USD/kWh}$$

$$C_{\text{unit}} = 0.091 \times 1.34 = 0.122 \text{ CDN/kWh}$$

- Process steam

$$m = 1.25 \text{ kg/s} \quad p = 1 \text{ bar}$$

$$a = 0.000023 + m^{(-0.1)} = 0.000023 \times 1.25^{(0.1)} = 0.0001881519$$

$$b = 0.0034 \times p^{0.05} = 0.0034 \times 1^{0.05} = 0.0034$$

$$C_{\text{unit}} = 0.0001881519 \times 620 + 1.045 \times 0.0034 = 0.015 \text{ USD/kg}$$

$$C_{\text{unit}} = 0.015 \times 1.34 = 0.020 \text{ CDN/kg}$$

- Demineralized water

$$m = 0.032389 \text{ m}^3/\text{s}$$

$$a = 0.005 + 0.0002 \times m^{(-0.6)} = 0.005 + 0.0002 \times 0.032389^{(0.6)}$$

$$= 0.006566$$

$$b = 0.0034$$

$$C_{\text{unit}} = 0.006566 \times 620 + 1.045 \times 0.0034 = 4.075 \text{ USD/m}^3$$

$$C_{\text{unit}} = 4.075 \times 1.34 = 5.460 \text{ CDN/m}^3$$



- toxic or hazardous solids & liquids (grass-root plant)

$$a = 2 \times 10^{-3} \quad b = N/A$$

$$\begin{aligned}\text{Unit cost} &= a \times \text{CE plant cost index} + b \quad (\text{USD/kg}) \\ &= 2 \times 10^{-3} \times 620 \quad \text{USD/kg} \\ &= 1.24 \quad \text{USD/kg}\end{aligned}$$

$$\text{Unit cost} = 1.24 \text{ USD/kg} \times 1.34 \text{ CAD/USD} = 1.66 \text{ CAD/kg}$$

I.8.3 Utility Cost Sample Calculation



Utility cost - Process Steam

- from class V estimation, the annual consumption of process steam from the sum of Ammonium Treat plant, Hydrogen plant, & sulphur plant was 715,930,418 kg/yr

$$\text{Annual Cost} = 715,930,418 \text{ kg/yr} \times 0.020 \text{ CAD/kg} = \boxed{14,318,608 \text{ CAD/yr}}$$

Utility cost - Demineralized Water

- Only cooler C-A01 / C-B01 consumes demineralized water and the consumption rate is 97,265.92 kg/h

$$\begin{aligned} \text{- Annual consumption} &= 97,265.92 \text{ kg/h} \times 24 \text{ h/d} \times 365 \text{ d/yr} \\ &\quad \times 0.9 \\ &= 766,844,493.4 \text{ kg/yr} \end{aligned}$$

$$\begin{aligned} \text{Annual consumption} &= 766,844,493.4 \text{ kg/yr} / 997 \text{ kg/m}^3 \\ &= 769,151.9492 \text{ m}^3/\text{yr} \end{aligned}$$

$$\text{- Annual Cost} = 769,151.9492 \text{ m}^3/\text{yr} \times 5.46 \text{ CAD/m}^3 = \boxed{4,260,104.36 \text{ CAD/yr}}$$

Utility cost - Cooling water

- From class V estimation, only sulphur plant has a consumption of cooling water @ a annual flow of 2910884.844 m³/yr

$$\text{- Annual Cost} = 2910884.844 \text{ m}^3/\text{yr} \times 0.0665125 \text{ CAD/m}^3 = \boxed{193,610 \text{ CAD/yr}}$$



Utility cost - Wastewater Disposal

- S-52 is the wastewater that needs to be treated.
- Information from Stimulation are :
 - $W_t = 7029 \text{ kg/h}$
 - Since this water contains some ammonium salts, and the water pH is high, it belongs to the Toxic or hazardous solids and liquid in U&V. The unit price for disposal have been determined to be 1.66 CDN/kg
 - Annual Utility Cost :

$$\text{Cost} = 7029 \text{ kg/h} \times 24 \text{ h/d} \times 365 \text{ d/yr} \times 0.9 \times 1.66 \text{ CDN/kg}$$

$$= 91,991,615 \text{ CDN/yr}$$



Utility Costs - Electricity

ex) Pump P-02

According to Ulrich & Vasudevan (2004) Figure 4.2

At $W_s = 0.8467 \text{ kW}$, motor efficiency $\epsilon_d = 0.83$

$$\text{motor power} = \frac{W_s}{\epsilon_d} = \frac{0.8467 \text{ kW}}{0.83} = 1.020 \text{ kW}$$

$$(1.020 \text{ kW})(24 \text{ h})(365 \text{ days})(0.9) = 8042 \text{ kWh}$$

I.8.4 General Operating Expenses Sample Calculation



Diluent Cost Calculation

- A question to Natural Resources Canada (2020) is as following

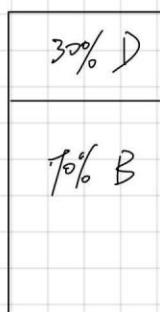
" Implied Bitumen Price: The implied bitumen price is derived from the price of Western Canadian Select (WCS), which is assumed to have a blend composition of 30% condensate and 70% bitumen. The value of the condensate is subtracted from the WCS price to arrive at the implied price of bitumen. The implied bitumen price represents the value of bitumen extracted from the oil sands before condensate is added to allow the oil to flow through pipelines. "

- According to Natural Resources Canada (2020), taking the average prices for WCS and Implied Bitumen the results are :

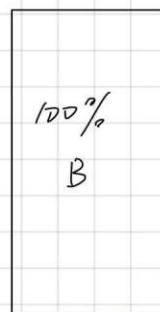
$$P_{WCS} = 325 \text{ C$D/m}^3 \quad P_{Bitumen} = 325.58 \text{ C$D/m}^3$$

- An illustration of how diluent (condensate) price is obtained is as follow :

$$D = \text{Diluent price (C$D/m}^3) \quad B = \text{Implied Bitumen price (C$D/m}^3)$$



WCS



Implied Bitumen

$$B \times 70\% + D \times 30\% = WCS$$

$$0.3 D = WCS - 0.7 B$$

$$D = \frac{WCS - 0.7 B}{0.3}$$

$$= \frac{325 - 0.7 \times 325.58}{0.3} = 323.64 \text{ C$D/m}^3$$

To Double check :

$$323.64 \times 30\% + 325.58 \times 70\% =$$

30% D
70% B

$$= 325 \text{ C$D/m}^3 \\ = \text{WCS price}$$



Operating Labor

- operating factor = 0.9

- According to Statistics Canada, Table 14-10-0326-02

the average offered hourly wage, for North American Industry Classification System : Petroleum and coal product manufacturing positions [324], the Statistics information are as follow:

Q1 2019	Q2 2019	Q3 2019	Q4 2019
No data	34.9	31.6	No data

- the average wage rate in 2019 is:

$$\frac{34.9 + 31.6}{2} = 33.25 \text{ CAD/hr}$$

- then the Annual wage is

$$33.25 \text{ CAD/hr} \times 24\text{hr/d} \times 365 \text{ d/yr} \times 0.9 = 262,143 \frac{\text{CAD}}{\text{yr}}$$

- According to Ulrich & Vasudevan (2004) p414.

Table 6.2 Operator Requirements for Various Types of Process Equipment.

Detailed numbers are listed in the Excel sheets,

The total # of op. labor = 5.75

$$\text{Annual Salary} = 262,143 \times 5.75 = 1,507,322.25 \text{ CAD/yr}$$



Class IV estimation - operating cost of Zinc Oxide Bed

- According to Hassan, Khannas, and Al-Mayah (2008) the H₂S removal process usually use 90 wt% ZnO + 10 wt% Alumina.

The efficiency is usually 90%.

- The market price, according to a online trading site, Alibaba (2020), are

\$2000 USD/t for ZnO powder from a supplier from Shanghai

\$15 USD/kg for Alumina powder, in 2020.

- Assume 90 wt% ZnO and 10 wt% Alumina, catalyst price is:

$$P_{cata.} = 2000 \text{ USD/t} \times \frac{1t}{1000kg} \times 90\% + \\ 15 \text{ USD/kg} \times 10\% = 3.3 \text{ USD/kg}$$

$$P_{cata.} = 3.3 \text{ USD/kg} \times 1.34 \text{ CAD/USD} \\ = 4.42 \text{ CAD/kg}$$

- From simulation, the feed to Zinc Oxide Bed is Stream S-29.

For S-29

$$F = 9310.5 \text{ kmol/h}$$

Sulfur molar fraction = 0.0654

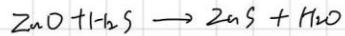
$$F_{H_2S, tot} = 9310.5 \text{ kmol/h} \times 0.0654 = 608.9 \text{ kmol/h}$$

- It's assumed that 99% of the H₂S will be processed by Amine Unit connected right next to stream S-29. The rest 1% of H₂S is being removed by Zinc Oxide Bed.

$$\Rightarrow F_{H_2S} = F_{H_2S, tot} \times 1\% = 6.089 \text{ kmol/h}$$



- The reaction is :



$$\Rightarrow F_{\text{ZnS}} = F_{\text{H}_2\text{S}} = 6.089 \text{ kwh/h}$$

- Annual consumption of ZnS is (Molar weight of ZnS is 97.474 kg/kmol)

$$m_{\text{ZnS}} = 6.089 \text{ kwh/h} \times 97.474 \text{ kg/kmol} \times 24 \times 365 \times 0.9$$

$$= 4679305.26 \text{ kg/yr}$$

- For a 90 wt% ZnO 10 wt% Alumina (Molar mass 101.96 kg/kmol)

$$m_{\text{Al}_2\text{O}_3} = \frac{m_{\text{ZnO}}}{90\%} \times 10\% = \frac{4679305.26 \text{ kg/yr}}{0.9} \times 0.1 = 519922.81 \text{ kg/yr}$$

$$\Rightarrow m_{\text{ore}} = m_{\text{ZnO}} + m_{\text{Al}_2\text{O}_3} = 519922.81 \text{ kg/yr}$$

\Rightarrow Annual Cost

$$\begin{aligned} C_{\text{ore}} &= m_{\text{ore}} \cdot P_{\text{ore}} \\ &= 519922.81 \text{ kg/yr} \times 4.42 \text{ CDN/yr} \\ &= 22980.588 \text{ CDN/yr} \end{aligned}$$

- For annual solid waste disposal, it had been determined to be

$$1.66 \text{ CDN/kg}$$

\Rightarrow Cost for solid waste disposal is

$$\begin{aligned} C_{\text{solid waste}} &= 519922.81 \text{ kg/yr} \times 1.66 \text{ CDN/kg} \\ &= 8630718.59 \text{ CDN/yr} \end{aligned}$$

\Rightarrow The total operating cost for Zinc Oxide Bed is

$$C_{\text{ore}} + C_{\text{solid waste}} = 22980.588 + 8630718.59$$

$$= \boxed{81,611,306 \text{ CDN/yr}}$$



Specific Gravity Sample Calculation

- According to Gray, M.R. (2015) P.15, Equation 1.1.

$$\text{oAPI} = \frac{1.415 \times 10^3}{\rho} - 131.5$$

- The product density is 736.3 kg/m^3

$$\Rightarrow \text{oAPI} = \frac{1.415 \times 10^3}{736.3} - 131.5 = 60.7$$



payout time

- According to CIE 664 Notes, the payout time is defined as

$$POT = \frac{\text{Total Capital Cost}}{\text{Gross Material Income}}$$

where Gross Material Income = Revenue - operating expense

$$- POT = \frac{6913450594}{4018036780 - 4011102003} = \boxed{9.7 \text{ years}}$$

→ If the product price rise 20% (WTI Cushing price)

Then Revenue becomes \$4631775286

$$POT = \frac{6913450594}{4631775286 - 4611102003} = \boxed{11.1 \text{ years}}$$

I.8.5 Discounted Cash Flow Sample Calculation



Sample Cash Flow for the First Operating Year (2025)

- Calculated results for the year 2020

$$\text{Total Fixed Capital} = C_{FC,20} = 6913450594 \text{ CDN}$$

$$\text{Working Capital} = C_{WC,20} = 1037017589 \text{ CDN}$$

$$\text{Manufacturing Expense} = C_{ME,20} = 401102003 \text{ CDN}$$

$$\text{Total Revenue} = R_{20} = 4018036780 \text{ CDN}$$

$$\text{Tax Rate} = r_t = 27\%$$

$$\text{Discount Rate} = r_d = 15\%$$

$$\text{Inflation Rate} = r_I = 3\%$$

- Manufacturing Expansion in 2025 (inflation for 5 years)

$$C_{ME,25} = C_{ME,20} \times (1 + r_I)^5 =$$

$$(-401102003)(1+0.03)^5 = -4649966561 \text{ CDN}$$

- Revenue in 2025 (inflation for 5 years)

$$R_{25} = R_{20} (1 + r_I)^5 = 4018036780 \times (1+0.03)^5 \\ = 4658005869 \text{ CDN}$$

- Total Capital Investment & Working capital in 2025
Since all the investments are done in the first five years

$$C_{FC,25} = 0 \quad C_{WC,25} = 0$$

- Cash Flow Before tax in 2025

$$= (-C_{FC,25}) + (-C_{WC,25}) + R_{25} + (-C_{ME,25})$$

$$= 0 + 0 + 4658005869 + (-4649966561)$$

$$= 8039308 \text{ CDN}$$

- Capital Cost Allowance and Taxes in 2025.

CCA for the first operating year, 2025, is 15%

The un-depreciated capital from last year, 2024 is
6913450594 CDN



$$CCA_{25} (15\%) = 6913450594 \times 15\% = 1037017589 \text{ CDN}$$

The depreciated capital is then

$$6913450594 - 1037017589 = 5876433005 \text{ CDN}$$

- Taxable income for 2025 = Cash flow 2025 - CCA₂₅
 $= 8039307 - 1037017589$
 $= (-1028978281)$

$$\text{Taxes} = (-1028978281) \times 27\% = (-277824136)$$

- Since the cash flow is negative, the taxes here becomes a credit.

- Cash flow after tax = Cash flow before tax in 2025 - tax
 $= 8039308 - (-277824136)$
 $= 285863444$

- Discount factor in 2025 = $(1 + r_D)^{-5} = (1 + 0.15)^{-5}$
 $= 0.297177$

- Discounted cash flow = Cash flow after tax × Discount factor
 $= 285863444 \times 0.297177$
 $= 142124654$

- Cumulative Discounted Cash Flow
 $= \text{Discounted cash flow in 2024} + \text{Discounted cash flow in 2025}$
 $= (-537113827) + 142124653$
 $= (-5233989174) \text{ CDN}$

- Note the amount of cumulative cash flow from 2024 of (-537113827) is also obtained in a same manner.
- the same calculation applies to rest of the years

I.9 REFERENCES

CH E 464 and CH E 465 class notes and tutorials.

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Gray, M.R. (2015). “Upgrading oilsands bitumen and heavy oil”, 1st Ed., The University of Alberta Press.

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SLURRY HYDROCRACKER PROJECT

Appendix J - Technology Evaluation

PREPARED FOR

Frank Nolte, PEng

Worley Limited

PREPARED BY

Team 15: TR Solutions

Jaryl Schmidt, Student

Jose Te Eng Fo, Student

Naira Correia, Student

Xingming Shan, Student

Yichun Zhang, Student

DATE: April 9th, 2020

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J.1 SUMMARY

This appendix contains the technology evaluation matrices for the catalyst selection and hydrogen production.

J.2 TECHNOLOGY EVALUATION MATRIX

Table J1. Process catalyst evaluation [1, 2, 3, 4].

Catalyst	Heterogeneous (Low-Activity) Catalyst			Homogeneous (High-Activity) Catalyst		
	Score	Weight Factor	Weighted Ranking	Score	Weight Factor	Weighted Ranking
Amount of Catalyst Required	3	6	18	6	6	36
Catalyst Cost	6	9	54	3	9	27
Rate of Desulfurization	6	6	36	9	6	54
Process Complexity	9	6	54	1	6	6
Total	162			123		

Table J2. Hydrogen production evaluation [5, 6, 7].

Hydrogen Production Technology	Steam Methane Reforming			Partial Oxidation of Heavy Hydrocarbons		
	Score	Weight Factor	Weighted Ranking	Score	Weight Factor	Weighted Ranking
Hydrogen Purity	9	6	36	9	6	36
Capital Cost	6	3	18	3	3	9
CO ₂ Emissions	6	3	18	6	3	18
Process Efficiency	6	6	36	3	6	18
Total	108			81		

Score:

- 1 - Least Desirable Option
- 3 - Less Desirable Option
- 6 - More Desirable Option
- 9 - Most Desirable Option

Weight Factor:

- 1 - Low Importance
- 3 - Medium-Low Importance
- 6 - Medium-High Importance
- 9 - High Importance

J.3 REFERENCES

- [1] Speight, J.G. (2017). “Handbook of petroleum refining”, CRC Press, Taylor & Francis Group
- [2] Zhang, S., D. Liu, W. Deng, and G. Que, “A Review of Slurry-Phase Hydrocracking Heavy Oil Technology”, Energy Fuels **21**, 3057-3062 (2007).

- [3] Sahu, R., B.J. Song, J.S. Im, Y.P. Jeon, and C.W. Lee, "A review of recent advances in catalytic hydrocracking of heavy residues", *J IND ENG CHEM* **27**, 12-24 (2015).
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- [7] Topsoe Technologies. (n.d.). *Large-scale Hydrogen Production*. Retrieved from https://www.topsoe.com/sites/default/files/topsoe_large_scale_hydrogen_produc.pdf

SLURRY HYDROCRACKER PROJECT

Appendix K - Email Correspondence with Mr.
Frank Nolte

PREPARED FOR

Frank Nolte, PEng

Worley Limited

PREPARED BY

Team 15: TR Solutions

Jaryl Schmidt, Student

Jose Te Eng Fo, Student

Naira Correia, Student

Xingming Shan, Student

Yichun Zhang, Student

DATE: April 9th, 2020

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K.1 SUMMARY

This appendix shows all the correspondence emails between the industrial advisor and the team, as well as, the meeting minutes for the three industrial advisor meetings.

K.2 INITIAL PROJECT DESCRIPTION

Project Design Basis Statement: CH E 435/465 Design II

Project Title: Bitumen Conversion Using Slurry Hydrocracking

Description: Most bitumen conversion in Alberta is achieved by carbon rejection processes such as delayed coking or ebullated bed hydrocracking (EB) (H---Oil or LC---Fining). More Recently Slurry Hydrocracking (SHC) processes are emerging that promise higher conversion of bitumen to synthetic crude oil. This project will develop economic basis for an SHC based bitumen upgrader

Strategic Intent: SHC promises higher conversion of bitumen into saleable products, which improves the overall project economics and may decrease the environmental footprint

Deliverables: The project will develop a concept study the provides the basis, heat and material balance, preliminary equipment sizing and cost for a 100,000 barrel per day bitumen upgrader the uses an SHC unit for bitumen conversion. The objective is to determine anticipated return on capital for this form of bitumen upgrading. To do this the team will prepare the following deliverables:

1. A Project and team management plan
2. Research options, data acquisition, and evaluation of options to determine project scope
3. BFD & PFD development and process simulation/modeling, with material and energy balances
4. Equipment selection and specification (preliminary) P&ID for a single specific piece of equipment
5. Analysis of regulatory needs: safety, environment, economic and applicable code and standards issues and requirements
6. Safety and Risk Management including process PFD PHA and P&ID Hazard and operability analyses
7. Capital and operating cost estimates
8. Economic and sustainability project evaluations
9. Preliminary project execution strategy
10. Technical written and oral presentations of deliverables

Problem Statement: This project includes both a reaction (hydrocracking) and product separation.

Primary Objective: The primary objective is to determine the return on capital of a standalone merchant upgrader using slurry hydrocracking to convert the bitumen into a pipeline---able synthetic crude

Feedstock Design Basis: The feedstock to the upgrader is assumed to be 100,000 barrel per day of hot bitumen product from a bitumen producer (typical steam---assisted gravity drainage production site in Alberta)

Product Design Basis: The synthetic crude from the upgrader will essentially be the liquid product from the SHC unit. Sales price of the final product will depend on the yield of various synthetic crude components (naphtha 50--170 C boiling range, light gas oil 170---350 C boiling range and heavy gasoil 350---550 C boiling range). The team will determine the value of the end product based on its composition.

Additional Information (where team research fails to find adequate info it will be provided).

K.3 EMAIL CORRESPONDANCE

On Tue, Jan 14, 2020 at 5:08 PM, Naira Correia <nairamar@ualberta.ca> wrote:

Dear Frank Nolte,

My name is Naira Correia and I am the leader for the capstone team working on the Slurry Hydrocracker project. CC'ed are my colleagues Jose Te Eng Fo, Jaryl Schmidt, Yichun Zhang, and Xingming Shan.

We are excited and looking forward to working with you for the next upcoming weeks.

Kind regards

Naira Correia (and colleagues)

On Tue, Jan 14, 2020 at 5:10 PM, Nolte, Frank (Edmonton) <frank.nolte@advisian.com> wrote:

Hi Naira

Hello to you and your team. This hopefully will be an interesting project for you and your team and an overall good learning experience.

And as it's -30C outside you have nothing to distract you from getting started!

Look forward to our first meeting.

Regards

Frank Nolte

Principal Consultant
Global Downstream and Market Services

On Fri, Jan 17, 2020 at 1:23 PM, Naira Correia <nairamar@ualberta.ca> wrote:

Dear Frank Nolte Peng,

It's Naira from TR Solutions,

We are currently trying to create a more detailed scope for the project, and some questions have arisen from our research. From this, we were hoping you could provide some clarity to help us solve some of the uncertainties we are currently facing.

The questions we currently have are:

- What is the intended location of our site, and what is the location factor?
- How would you like us to price our product?
- What are the hurdle and discount rates?
- What is the currency you would like us to use, and what are the time-bases?

Additionally, we were wondering if you had any recommendations for source material that we can use?

We are looking forward to your reply,

Thank you for your time and patience,
Naira and the rest of the team

On Mon, Jan 20, 2020 at 11:33 AM, Nolte, Frank (Edmonton) <frank.nolte@advisian.com> wrote:

Hi 465 Team

Hope that you have made a good start on your research related to this topic. A few answers to point you in the right direction.

- What is the intended location of our site, and what is the location factor? **Plant located in Edmonton area YEG vs USGC factor is 1.35**
- How would you like us to price our product? **The product is priced a slight discount to the West Texas Intermediate (WTI) benchmark = Use province of Alberta Petroleum Marketing Commission price projections for WTI**
- What are the hurdle and discount rates? **I would use a hurdle rate of 15% for this project**
- What is the currency you would like us to use, and what are the time-bases? **Use Cdn\$ the project would take 5 years to develop and come on stream in 2026. Base the project life at 35 years from Startup**

Additionally, we were wondering if you had any recommendations for source material that we can use? **Look up slurry phase hydrocracking see if you can find information from the companies UOP and CLG also look for CanMet pilot test work done at Petro-Canada Edmonton Refinery**

Frank Nolte
Principal Consultant
Global Downstream and Market Services

On Wed, Jan 22, 2020 at 09:31 AM, Nolte, Frank (Edmonton) <frank.nolte@advisian.com> wrote:

Team

You may find this useful in your research of slurry hydrocracking technology.

Frank Nolte

Principal Consultant
Global Downstream and Market Services



**Honeywell
uop**

REFINING

**UOP Uniflex™ MC™
Residue Hydrocracking Process**

Robust, cost-effective solution for residue conversion

Introduction

The Uniflex MC process converts vacuum residue and other heavy feedstocks into higher-valued distillable products. This high conversion slurry hydrocracking technology combines elements of several commercial technologies: CANMET™ Hydrocracking process, UOP Unicracking™ process, UOP Unionfining™ process, and employs highactivity molybdenum based MicroCat™ catalyst. The catalyst is produced on site from readily available commodity chemicals. Uniflex MC is easily integrated into existing refinery configurations, providing refiners the ability to increase bottom-of-the-barrel conversion, reduce residual fuel oil production and increase production of light fuels. The product streams are also suitable as high value petrochemical complex feedstock.

Technology Delivery

Uniflex MC is delivered as two separate technology packages: the Uniflex Process and the MC Catalyst System. Both are delivered through UOP's 'Schedule A' Basic Engineering design package. The MC Catalyst System can also be supplied as part of an engineered, modular supply from UOP's Modular Equipment Group to reduce on-site construction and facilitate enhanced process automation and advanced control.

Process Chemistry

The Uniflex Process unit is a thermal cracking / hydrogenation process. Residue molecules are thermally cracked by application of heat at high pressure and with sufficient residence time. The cracked molecules react with hydrogen in the presence of catalyst to produce stable, lighter products predominantly in the transportation fuel range. As a side reaction, free radicals produced as a result of thermal cracking can react to form even larger free radicals which form a "mesophase" of condensed materials. The catalyst inhibits this reaction by terminating the free radicals. A second function of the catalyst is to prevent any mesophase formed from coalescing and forming larger molecules. The relationship between operating temperature, reactor size and catalyst addition rate can be optimized to a refiner's particular processing and economic situation.

Process Description

The configuration of the Uniflex process is very similar to that of a conventional hydrocracking process, consisting of a reaction section, a separation section to recover liquid and gaseous reaction products, and a fractionation section to separate its product into the various boiling-range fractions required by a refiner.

Feedstock, combined with catalyst, and hydrogen-rich recycle gas are heated and then introduced to the bottom of the first upflow reactor at operating temperature.

FEATURES & BENEFITS

High Reliability Design

- Based on a proven reactor system based on a commercial unit that demonstrated over 15 years operation.
- Supported by 30 years of extensive research, pilot plant operations, and development.
- UOP's expertise in high-pressure hydroprocessing which is reflected in the design, technical services, and project optimization capabilities associated with Uniflex Technology.

Ease of Operation

Generates the Highest Product Value

Uniflex has the highest total yield of valued products with 95-98% conversion of a wide range of low-value residue feedstocks. Selectivity to distillate products is the hallmark of Uniflex, with production of LPG, Naptha, and Diesel as fuels or petrochemical feedstock.

High Product Quality as Products or Feedstocks

The integrated Uniflex/DHT process produces reformer grade Naptha, Euro V Diesel and sweet, olefin-free LPG. The LPG and Light Naptha are excellent Steam Cracker feeds with high normal paraffin content. LPG can be processed in an FCC or Hydrocracking unit, with or without pretreatment in the Uniflex unit.

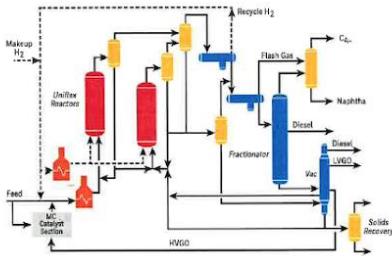
Integration with Other Units

Due to the number of products produced by any slurry hydrocracking technology, integration with the existing or new complex is critical to extract full value from the technology. UOP is uniquely positioned to optimize a total project with a broad understanding of refining and petrochemical configurations.

The reactor's unique design promotes intensive back-mixing of catalyst and reactants without reactor internals. The resulting near-isothermal conditions allow the reactor to operate stably at the higher severity required to achieve maximum residue conversion, while minimizing undesirable secondary cracking reactions that produce lower-valued gaseous byproducts.

The first reactor's liquid effluent is routed to the bottom of the second reactor. The second reactor effluent and vapor from the first reactor flow to the unit's separation section where liquid and vapor fractions are recovered. Vapors are scrubbed, combined with makeup hydrogen and recycled back to the reaction section. Liquid from the separator section flows to the unit's fractionation section for product recovery. A portion of the unconverted residue (pitch) is recycled to the second reactor. The rest of the unconverted residue can be used as a low-cost substitution fuel for conventional heavy liquid residue or coker feed, or used as a solid fuel for end users such as fluidized boilers and gasifiers. Optionally, the pitch can be sent to a Solids Recovery section to produce additional liquid products and a solids stream, which can be sent for metals recovery.

Uniflex MC Process Configuration



Additional Processing Options

- Integrated Distillate Hydrotreating: With the addition of six pieces of equipment fuels products can be treated to produce sweet LPG, reformer grade Naphtha and Euro V quality Diesel. This is at a cost of about 50% of a new stand-alone DHT unit.
- Enhanced LPG Recovery: Where LPG has a higher value than fuel gas, recovery of this product can be increased to 99% at an effective capital cost and low utility consumption.
- Side-on VGO Hydrotreating: VGO quality can be cost effectively improved with the addition of a small HT reactor on the product stream.

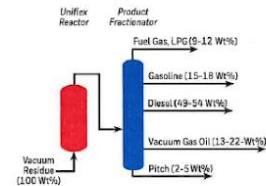
Product Yields

The yield structures produced when processing Vacuum Residue from a typical sour crude in a Uniflex process unit are illustrated in the figure shown below. The advantages of the yields versus other technologies include:

- Ability to be highly selective to Naphtha and Diesel
- Flexibility to shift to heavier products if desired
- Total liquid yields up to 115 vol%

These advantages provide refiners the ability to more closely tailor their operations to satisfy projected declining residual fuel oil markets and increasing fuel and petrochemical feedstock demands.

Uniflex Process Typical Yields



MicroCat Catalyst

The high-activity catalyst (MicroCat) used in the Uniflex process is generated on-site in the MC Catalyst Section. The refiner will produce the MicroCat from commodity raw materials readily available in bulk quantities on the open market. This section contains simple process operations to produce consistent, high quality catalyst. On site quality assurance testing is part of the technology package.

Additional Opportunities

Refiners can also benefit from the synergies which arise when the Uniflex MC process is used in association with other Refining and Petrochemical processes. The Uniflex MC process unit can be used to upgrade/convert marginal-quality refinery streams such as FCC slurry oils and heavy coker gas oils. For refiners with existing delayed coking units who are considering increasing refinery throughput, the addition of a Uniflex process can debottleneck the delayed coking unit and provide a significant increase in the refinery's overall liquid product yield.

For more information

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On Wed, Jan 22, 2020 at 2:20 PM, Naira Correia <nairamar@ualberta.ca> wrote:

Dear Nolte,

Thank you for addressing our concerns we really appreciate you taking your time to respond to us. We are looking forward to meeting you on Tuesday!

Thank you,

Naira Correia and the Team

On Mon, Jan 27, 2020 at 08:32 AM, Nolte, Frank (Edmonton) <frank.nolte@advisian.com> wrote:

Hi Team

Thanks for a very comprehensive situation report. You are clearly off to a great start on this work.

Prior to the meeting today, I'll give you a few answers to your questions, which we can discuss in more detail in our meeting today. One thing to note, you will run sensitivities on your economics to test your assumptions for variances of say 10 and 20%. This will give you some idea of the risk profile associated with each assumption you've made on feedstock pricing, product pricing and capital cost

See you all later. Looking forward to meeting the team.

Regards

1. Pricing Scheme for Crude Product
 - (BP cuts, Sulfur, Density) **You will find pricing of crude oil by density (API gravity) and sulphur content online assume theses**
2. What is the reference price for natural gas and Athabasca bitumen on the date Jan 7th, 2020?

Natural gas price : use <https://www.statista.com/statistics/252791/natural-gas-prices/>

- How do we estimate the price of bitumen over the 35 years? **Assume the same discounted price over WTI for the period**
- 3. Is the assumption of the Athabasca bitumen an acceptable assumption? **Yes**
- 4. What is the industrial price for an iron-based catalyst (eg. FeSO4)? **This catalyst is cheap (assume \$5.00/lb inflated at your assumed inflation rate off to the end of the project)**
- 5. Crude product will be assumed to be the same as CANMET product. Is that an acceptable assumption for the business case. **What yield for each component (boiling range) did you assume?**
- 6. Do you have any suggestions for what to do with the sulfur product?
 - Sell it? Stockpile it? **Look up global sulphur price and assume that value projected out into the future.**

Frank Nolte
Principal Consultant
Global Downstream and Market Services

On Tue, Feb 11, 2020 at 11:45 AM, Naira Correia <nairamar@ualberta.ca> wrote:

Good morning Mr. Nolte,

At the moment we are working on simulation and sizing and we discovered that there is no kinetics information available for the iron-based catalyst (FeSO₄). Upon talking with professor Semagina and our academic advisor Arno de Klerk we came to two conclusions.

1- We need you to provide the CANMET reactor sizing specification (diameter and height), or can you provide any good reactor sizing estimates to calculate the space velocity for this specific catalyst.

2- The other option would change of catalyst which there is a lot of literature and research available.

The second option is very undesirable because it would require us to design a catalyst regeneration unit.

Best regards,

Naira and Team 15

On Tue, Feb 11, 2020 at 12:43 PM, Nolte, Frank (Edmonton) <frank.nolte@advisan.com> wrote:

Hi TR Team

Sorry for the delayed response. I was out of town this weekend.

LHSV for the reactor is .21/hr. The estimated vapor voidage (volume not occupied by liquid) in the reactor is 40%

Just as a check on the reactor size I have an equipment list for a 41,000 BPSD SHC unit that shows a reactor train consisting of 4 parallel reactors each 13.5 ft OD and 81 ft tangent to tangent length. With a design pressure of 2340 psig and design temp of 900 F. The reactor shell is constructed of 2 ¼ Cr 1Mo V mod stainless steel with a 3 mm TP 321/347 SS liner. The high-pressure portion of the SHC is all constructed of this same material.

I would like for the team to calculate the cross-sectional area and height of reactor required for your reactor based on the LHSV and voidage as a comparison to the datasheet numbers I have for the commercial unit that I have.

Hopefully that helps

Regards

Frank Nolte

Principal Consultant

Global Downstream and Market Services

On Sun, Mar 1, 2020 at 2:01 PM, Naira Correia <nairamar@ualberta.ca> wrote:

Good afternoon Mr. Nolte,

On behalf of TR Solutions (Team 15), I am glad to announce that our second meeting will be held on March 3rd at 2 pm on DICE 8A-246. Please see attached for the meeting agenda, list of questions for discussion and the situational report.

Please let me know if you have any questions and concerns,

See you on Tuesday,

Naira Correia and TR Solutions

On Sun, Mar 1, 2020 at 2:39 PM, Nolte, Frank (Edmonton) <frank.nolte@advisian.com> wrote:

Hopefully this helps for now. Saw you all on Tuesday.

Frank

1. For the sensitivity analysis are there any specific variables that we should consider? For the change in NPV would like 10% or higher?
 - You should vary the followig:
 - CAPEX. -20/+50%
 - OPEX -20/+20%
 - Cost of Capital – we should discuss your assumptions
2. What legislation do you want us to refer to for environmental analysis/ API Codes, ASME Codes? Refer to Alberta Environment Sulphur limited based on sulphur plant throughput, Estimate the GHG (CO₂ emissions).
3. Capacity (Flow Rates) is very high, making our sizing estimations grossly large. Are we able to split the process into two different trains? Certainly
4. How do we estimate steam flow rates from the sulfur recovery plant (WHB)? Check out <https://www.ogrtt.com/files/publications/256/2018-11%20Sulphur%202018%20-%20Factors%20Affecting%20Claus%20Waste%20Heat%20Boiler%20Design%20and%20Operation.pdf>

On Wed, Mar 15, 2020 at 2:51 PM, Naira Correia <nairamar@ualberta.ca> wrote:

Good Afternoon Frank,

We want to confirm a design consideration we implemented for our hydrogen compression train for the hydrogen supply. In traditional compressor design, a knockout drum is installed after intercoolers to knock out condensables to protect the compressor. The compressor train in our design is downstream of a pressure swing adsorber sending hydrogen stream with high purity with negligible to zero amount condensables. Thus, we decided to remove knock out drums but kept the intercoolers to maintain low temperatures.

Also, during the sizing of the three-phase separator D-03 the terminal velocity was found to be 0.47 m/s. With a flow rate of 0.7373 m³/s of mixture fed to the separator. The result:

Separator Minimum Diameter= 1m

Separator Minimum Length= 1.995m

Considering the scale of our separator, we think it might be too small. So alternatively, we want to have something larger, to handle different situations. In the same literature, the suggested L/D ratio was 2 to 4. Can we use a D=3m, and L=12m, with an L/D ratio of 4 instead?

Please let us know if this is a considerable option or if you have any other concerns.

Kindest regards,

Naira Correia

On Wed, Mar 15, 2020 at 6:37 PM, Nolte, Frank (Edmonton) <frank.nolte@advisian.com> wrote:

Hi Naira

Some. Short answers to your questions;

The interstate separator is often needed to remove small amounts of lube oil.

For the separator you also want to consider that you need space for 5 min of residence from the high and low liquid level to both the high high and lowlow liquid level. That also for operator response in the event of alarm

Sent from my Bell Samsung device over Canada's largest network.

On Sun, Mar 29, 2020 at 12:59 PM, Naira Correia <nairamar@ualberta.ca> wrote:

Good afternoon Mr. Nolte,

On behalf of TR Solutions (Team 15), I am glad to announce that our third and final meeting will be held on March 31st at 4 pm online via Zoom Meetings. Please see attached for the meeting agenda, final presentation slides and list of questions for discussion.

Please let me know if you have any questions and concerns,

See you on Tuesday,

Naira Correia and TR Solutions

K.4 MEETING MINUTES

Meeting #1 Minutes summary

Feedbacks from Mr. Frank:

- Consider the components nitrogen and oxygen in bitumen feed
Potential hydrogen consumers:
 $O \rightarrow H_2O$
 $N \rightarrow \text{amine}$
- Propane and butane are salable products, can improve the economics of the project
- Vacuum gas oil needs to be cracked further to be sold \rightarrow gasoline and diesel
- What to do with residue?
 - If you concentrate spent catalyst, think about what you can do with that stream...
 - Economic: sensitivity analysis on the cost of bitumen
 - Contact Marnie: for future pricing (Diluted Bitumen): Sold as WCS (Western Canadian Select) - work backward from that to get bitumen pricing. WCS Indexed on WTI...
 - Our product will be a premium compared so SSP made by Suncor/Syncrude/CNRL due to product distribution - very high distillate content in our product
 - Sidecar Hydrotreating?
 - When heating the bitumen: start stripping bitumen: as the temperature increases, start to allow large bitumen molecules to agglomerate - the precursor to coke formation
 - Want heavy end of gas oil for recycle to keep solvency during thermal cracking
 - Sequestration of SMR unit (likely outside the scope of this project)? Other Carbon Capture technologies \rightarrow liquify \rightarrow pump? Include a discussion about the CO₂ sequestration (how we can improve on CO₂ emissions)?

Overall: Mr. Frank said that we had a comprehensive situation report, he agreed with our proposed solution and made suggestions on what to be improved. The format of the meeting was very interactive, and all the team members had an opportunity to share their questions with Mr. Frank.

Meeting #2 Minutes summary

Meeting 2 was held on March 3rd at 2 pm on DICE 8th floor, Room 8A-246. All team members were present and participated during the meeting. Naira and Jaryl were the main presenters and the entire team participated in the discussion. All the pre-meeting material was submitted on time on Sunday including Mass and Energy balances, PFD and list of questions. The Agenda was sent just a few minutes late but on Sunday as well. All team members completed the ITP Metrics prior to the deadline. Overall the team worked well to achieve the requirements for meeting 2, we receive good feedback from Mr. Nolte, and although we are a week behind we have a strategic plan going forward.

- Frank Nolte will be out of the country for the next meeting (final presentation), therefore the final meeting will be one week before the expected date.
- Possibility of CO₂ Sequestration from PSA (CO₂ capture)
 - As an opportunity, we could sell it and aid for our economic analysis
 - PSA purge steams seems to be pretty high
- Top-end reaction temperature: 780 F = 465 C
- Reactor pressure can be brought back to 17 Mpa; solution loss H₂ at high pressure
- Capacity of hydrogen plant seems too high
 - Can do a hydrogen recovery before putting hydrogen back into PSA
- The sulfur plant
 - The cost might be lower than actual by a factor of 2
 - Most possible of sulfur = 840 tons/day = 5% of bitumen feed
 - May have to apply tail gas unit and sulfur recovery
- Amine plant
 - Only C1, C2, and C3 with partial pressure = around 93 kpa
- Residue conversion
 - 95% conversion is a much more reasonable conversion of a bubble reactor
 - 5% of the unconverted residue from the 665516kg/h still a lot, therefore, look for what opportunities and challenges are available when dealing with it
 - How much of the cement could be used (make better sidewalks)
- Fractionation system should have zero naptha, distillate
- Sensitivity analysis:
 - Capital cost -20/+50%; Operating cost +- 20%; Discount rate: 10 to 20%

Meeting #3 Minutes summary

Meeting 3 was held on March 31st at 4 pm via zoom meetings online. All team members were present for the meeting, as well as, Frank Nolte and Arno De Klerk. The pre-meeting material was submitted on time on Sunday including meeting agenda, final presentation PowerPoint, and list of questions. The presentation participation divided in the following manner Jaryl covered the technology selected, Jose covered the explanation for the PFD, Naira covered the safety and risk analyses, Yichun did the environmental analysis and Shan did the economic analyses. All team members completed the ITP Metrics prior to the deadline. Overall the team worked well to achieve the requirements for meeting 3, we receive good feedback from Mr. Nolte he was pleased with the results presented and he also gave us a few implementations and adjustments to make for the final project.

Mass/Energy balance:

- Heating value of the PSA vent gas in terms of a utility heating stream (how it compares with natural gas)
- How much of the SMR load is reactor off-gas
- How much valuable product is lost in the residue stream
- Show a very simple version of “if I have 1kg of bitumen, __g goes to SMR, __ g goes to residue, etc.

Economics:

- Look into hydrogen cost per m³ (production)
 - Possibly buy the hydrogen from another company and save on the capital cost
 - \$3/1000 std ft³ (rough number from Frank)
- \$6B erected cost doesn't seem too off (from what he estimates)
- Look at the diluent recovery there is a possibility of selling at a higher price
- Current analysis (we've done) is a light/heavy spread - not a bad metric
- CO₂ sequestration credit
- Make sure capital costs make sense