# AREA 3 DESIGN SUMMARY

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## Design problem definition

- Number and size of reactors

A total of 20 reactors will be used, each of bulk volume and working volume of 668.8 $m^3$ and 608 $m^3$, respectively. Liquefaction will be conducted in 4 reactors followed by saccharification in 16 reactors. This is to ensure that the production requirement of 71 900 kg $h^{-1}$ of glucose will be met.

- How to ensure sterility

The liquefaction reaction will be carried out at 90 $^oC$. The heating will be performed by low pressure steam, available at 131 $^oC$. The saccharification reaction will be carried out in a sterile environment created by steaming all the ports of the reactor with low pressure steam at 131 $^oC$ (P. M. Doran, 2012). In addition, all exhaust gas will be filtered from the reactor nozzles using monofilament polypropylene filters of appropriate pore sizes (S. Ramaswamy et al., 2013).

- Physical design of each unit operation within the flowsheet

The shell and tube heat exchanger heating towns water to 55 $^oC$, will be positioned at the entrance to Area 3 and the one cooling the resulting glucose solution to 35 $^oC$ will be positioned at the exit from Area 3. These units will be made of high carbon steel. The liquefaction reactors andsaccharification reactors, each 10.1 m high, will be made of grade 304 stainless steel. The liquefaction units will be positioned in a single row of four and the saccharification reactors in four rows of four. The first set of 4 centrifugal pumps will be positioned right after the liquefaction reactors and the second set of 4 will be positioned right after the saccharification reactors. Both sets will be made of high carbon steel.

- Physical design of any unit operation internals

The liquefaction reactor internals will consist of 4 baffles, a stainless-steel screen mesh at the bottom of each of the reactors and 2 six-blade turbines pitched at 45 $^o$ and mounted on a shaft. The saccharification reactors will also consist of 4 baffles and 4 six-blade turbines pitched at 45$^o$ and mounted on a shaft. In addition, internal cooling will be provided for the saccharification reactors by helical coils carrying cooling water at 18 $^oC$. A stainless-steel screen mesh will also be positioned at the bottom of the saccharification reactors. For both liquefaction and saccharification, all internal units of the reactors will be made of grade 304 stainless steel. The two shell and tube heat exchangers and 8 centrifugal pumps will have internals made of high carbon steel.

- <u>How to schedule the processes to minimise capital expenditure</u>

Liquefaction will be carried out in a total time of 2 h including charging and discharging time as well as downtime and reaction time. Saccharification will take place in 8 h under the same conditions. Four liquefaction reactors will be supplementing sixteen saccharification reactors. This will enable flexible transition to continuous operation if necessary as well as minimise the overall capital cost of reactors required to deliver 71 900 kg $h^{-1}$ of glucose.

- <u>Ensuring that unit operations are sufficiently optimised</u>

The liquefaction and saccharification reactors consist of agitation systems that function in turbulent flow and maximise mixing. In addition, baffles are in place that aid in eliminating the effects of vortices in each of the reactors (P. M. Doran, 2012). Moreover, internal cooling coils for the saccharification reactor enable more homogeneous heat transfer within the reactors and better temperature control (Perry et al., 2008).

- <u>Ensuring that unit operations can be controlled, and product specification maintained</u>

Appropriate control schemes are in place that control the flow of both the stream entering Area 3 from Area 2 as well as the stream of heated towns water. Flow ratio control and level control prevent liquefaction reactors from overfilling. In addition, temperature control of the saccharification reactors is provided in the form of cascade control to implement more rapid responses to disturbances in the flows of utilities being supplied (J. Love, 2008). Furthermore, control systems designed to monitor the pH and temperature of the saccharification reactors, allow maintenance of glucose specification to the desired concentration of 60 g $dm^{-3}$ at an overall production rate of 71 900 kg $h^{-1}$.

- <u>Consideration of how Areas interact and how they are started up and shut down</u>

Area 2 is a continuous process. There is a need for continuity in flow between Areas 2 and 3, without intermediate storage. During start up and shut down when flow is unsteady, additional control measures must be put in place between Areas 2 and 3. In addition, Area 4 is a batch process with a batch time of 7.5 h. Intermediate storage between Areas 3 and 4 is necessary since, the Area 3 batch time is 10 h. During start-up and shut-down, additional safety measures must be implemented between Areas 3 and 4.

<u>What additional unit operations are required and to ensure that sufficient design is undertaken</u>

Microfiltration membranes with high hydrolytic stability and pore size of 0.1 μm are required to ensure the complete purification of the glucose solution resulting from the saccharification reactors. These membrane modules must be made of polymeric material such as acrylic copolymers and are vital since their optimal pH is around 5 which is the same pH at which saccharification will be conducted (Perry et al.,2008).

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# 1 Process development

## 1.1 Area 3 sequence of operations

Area 3 commences with liquefaction of stream 3.1, as indicated on the PFD which can be found in section 5, appendix 5.1. This will take place in 4 stirred tank vessels with 30 minutes filling and 30 minutes emptying time for each vessel (L. G. Britton, 1999). Stream 3.2 will be heated to the desired temperature of 55 $^{o}$C in a shell and tube heat exchanger E-302, prior to being discharged into each of the 4 liquefaction reactors.

The new reactor system involving 4 liquefaction reactors followed by 16 saccharification reactors was devised to minimise idle time and ensure that in the case that continuous operation is desired this can also be made possible. The previous system of 10 batch lines, each having a liquefaction reactor and a saccharification reactor of volumes 239.7 and 958.7 $m^3$ respectively, would have wasted a lot of reactor volume, resulting in excess capital expenditure.

The new system consists of 20 reactors in total, each of bulk volume 668.8 $m^3$. Each of the four liquefaction reactors will be supplying 4 different saccharification reactors. This will be conducted in a controlled manner using a centrifugal pump E-303 along each route connecting one liquefaction reactor to 4 saccharification reactors.

Liquefaction will be performed at 90 $^{o}$C (Akerberg et al., 2000) and 1 bar as opposed to 1.8 bar, in a total time of 2 h including charging and discharging times. This change in operating pressure as opposed to the one stated in task B, considers head losses due to friction in pipework and allows for safer operation of the liquefaction process since, the operating temperature will be high.

The product of the semi-batch liquefaction process will then be discharged into the saccharification reactors, after removal of solid fibre residue which will be contained on the stainless-steel screen mesh positioned within each of the liquefaction reactors (A. K. Coker, 2007). This will be removed during down-time due to cleaning.

The saccharification reactors will be cooled down to 55 $^{o}$C using cooling water at 18 $^{o}$C and 5 bar. This temperature is the optimal temperature for the saccharification reaction (Akerberg et al.,2000).

Immobilised glucoamylase and α-amylase will be added, and these will be recycled at the end of each batch. Their recovery is made possible by the provision of a stainless-steel screen mesh at the bottom of each saccharification reactor. $H_2SO_{4\,(aq)}$ and $NaOH_{(aq)}$ will be available for pH control. The cooling water will be supplied through helical coils within each saccharification reactor and will also allow for additional cooldown due to the exothermic nature of the saccharification reaction (Y. B. Tewari et al., 1989).

Stream 3.6 resulting from the saccharification reactors will then be pumped by centrifugal pumps to a new pressure of 1.6 bar as indicated on the PFD. Stream 3.7 will then be cooled down to 35 $^{o}$C after passage through another shell and tube heat exchanger E-306. The cooling duty of E-306 will be provided by cooling water at 18 $^{o}$C.

The primary objective of Area 3 is to deliver 60 g $dm^{-3}$ of glucose solution to Area 4 for fermentation, at a production rate of 71 900 kg $h^{-1}$. The sequence of 4 liquefaction reactors followed by 16 saccharification reactors will fulfil this target since it will generate a glucose solution of 60 g $dm^{-3}$ in each saccharification reactor. This solution will subsequently be pumped to E-306 which is located at the exit of Area 3. However, the resultant solution will require further purification prior to entry to Area 4. An additional microfiltration unit is required to achieve this. This additional unit will need to have pore sizes of at least 100 nm and filter aids to reduce caking effects (Perry et al. 2008).

The major items of capital equipment of Area 3 include the 4 liquefaction reactors and 16 saccharification reactors, including their internal unit operations. These reactors are large in volume and overshadow the rest of the unit operations of Area 3 in terms of capital cost.

## 1.2 Description of Area 3 Control Scheme including alarms, start-up, shut-down and emergency shut-down procedures

A PCD of Area 3 can be found in section 5, appendix 5.2.

Control Scheme:

**E-301 -** Flow ratio control is employed by considering stream 3.1 coming from Area 2 to be wild. FIT-323 and FIT-333 measure the flowrates of both stream 3.1 and 3.3. The ratio of these flowrates is used to control CV-3 and as a result, adjust the flowrate of stream 3.1. In addition, a simple feedback control loop is employed to control the temperature in each of the liquefaction reactors. The simple feedback loop controls temperature by adjusting CV- 4 and prevents runaway side-reactions from occurring.

**E-302 -** A simple feedback control loop is employed to control the temperature of stream 3.3 entering each of the liquefaction reactors. The position of CV-1 on the utility line is adjusted. This mechanism manipulates the flow of the utility delivering the duty required by E-302 and ensures that stream 3.3 enters the liquefaction reactors at the right temperature.

**E-304 -** The flow of stream 3.5 entering the saccharification reactors is manipulated to control the level in each of the reactors. The position of CV-5 is regulated based on the level of liquid in each of the saccharification reactors measured by LIC-306. In addition, pH control is achieved in each of the saccharification reactors by use of a durable pH sensor made of inert polyphenylene sulfide. MIT-307 measures pH which is linearised by MIR-307 using the anti-logging approach (J. Love, 2008). This in turn is fed to MIC-307 which sends a split signal to both CV-7 and CV-8. The position of these valves is regulated to maintain pH at the desired set-point of 5. Moreover, cascade control is employed in controlling the temperature of the saccharification reactors. This is done to ensure that reaction temperature conditions do not deviate significantly due to disturbances in the flowrate of the cooling water utility being supplied (J. Love, 2008). TIT-309 and TIT-308 measure the temperatures of the cooling water utility and saccharification reactor, respectively. The output of TIC-309 is used to regulate CV-9.

**E-306 -** A simple feedback control loop is employed to control the temperature of stream 3.8 entering Area 4. The position of CV-10 on the utility line is adjusted. This mechanism manipulates the flow of the utility delivering the duty required by E-305, ensuring that stream 3.8 enters Area 4 at the right temperature of 35 $^{o}$C.

In addition, for scheduling operation between the liquefaction and saccharification reactors, a programmable logic controller is in place which monitors charging and discharging time of material within each reactor sequence.

Alarms:

A high-high level alarm and level switch are triggered in E-301, ceasing the flow of stream 3.3. As a result, flow-ratio control causes stream 3.1 to cease flowing.

Start-up procedures:

CV-1, CV-2, CV-3, and CV-4 open. The programmable logic controller sets the operation along each reactor sequence for charging and discharging time (30 minutes). CV-5 and all control valves related to E-304 open. CV-10 opens.

Shut down procedures:

CV-2 shuts. CV-3 shuts down. CV-1 and CV-10 shut down.

Emergency shut down procedures:

CV-5 shuts. The programmable logic controller ceases the operation along each reactor sequence. CV-3 and CV-2 shut. CV-1 and CV-10 shut down.

# 2 Saccharification reactor Piping and Instrumentation Diagram

A P&ID of the saccharification reactor can be found in section 5, appendix 5.3.

# 3 Size, power, and cost estimates of the process equipment in Area 3

## 3.1 Liquefaction and Saccharification Reactors

In the new reactor system consisting of 4 liquefaction reactors followed by 16 saccharification reactors, dimensions of key parameters associated with each reactor type are entirely different to those calculated in task B.

The working volume of 608 $m^3$ was estimated by assuming a charging and emptying time of 30 minutes for each reactor (L. G. Britton, 1999). The charging and discharging times are included in the assumption of 2 h and 8 h batch time for liquefaction and saccharification, respectively. The bulk volume of 668.8 $m^3$ was estimated by increasing the new working volume by 10 %. The new dimensions of each reactor type and its internals are listed in table 1.

*Table 1: New parameter dimensions of liquefaction and saccharification reactors*

| **Dimension** | **Reactor type** | |
|---|---|---|
| | **Liquefaction** | **Saccharification** |
| **Reactor height / m** | 10.10 | 10.10 |
| **Reactor diameter / m** | 9.18 | 9.18 |
| **Reactor wall thickness / m** | 0.0604 | 0.0604 |
| **Impeller diameter / m** | 3.03 | 3.03 |
| **Impeller blade width / m** | 0.606 | 0.606 |
| **Impeller blade thickness / m** | 0.003 | 0.003 |
| **Distance between impellers / m** | 4.11 | 1.37 |
| **Baffle length / m** | 6.32 | 6.32 |
| **Baffle width / m** | 0.918 | 0.918 |
| **Shaft length / m** | 9.10 | 9.10 |
| **Shaft diameter / m** | 0.562 | 0.562 |
| **Shaft thickness / m** | 0.004 | 0.004 |
| **Stainless steel screen length / m** | 6.95 | 7.91 |
| **LPS sparger length / m** | NA | 4.29 |

All relevant dimensions listed in table 1, were calculated under the same assumptions employed in task B. For liquefaction, 4 reactors are involved, each comprising of an external heating jacket and 2 internal six-blade pitched ($45^o$) turbines. For saccharification, 16 reactors are involved each comprising of internal cooling coils and 4 six-blade pitched ($45^o$) turbines.

Additional parameter values calculated that were not incorporated in the calculations performed in task B include, impeller blade thickness and baffle thickness. A blade thickness of 3 mm was assumed (R. Mezaki et al., 2000). Baffle thickness was estimated by dividing reactor diameter by 100 (R. Mezaki et al.,2000).

In addition, table 2 lists the key dimensions of the heating jacket for each liquefaction reactor and internal cooling coils for each saccharification reactor. These were estimated by specifying the area available for heat transfer at 182.4 $m^2$ and 64.2 $m^2$ for the liquefaction and saccharification reactors, respectively. This assumption carries with it a degree of uncertainty of approximately 40 %, since calculations were performed by using scale drawings of the reactors as well as assuming that the contents of the reactors are mainly water.

*Table 2: Dimensions of liquefaction reactor heating jacket and saccharification reactor internal helical coil cooling system*

| | **Heating jacket** | **Helical cooling coils** |
|---|---|---|
| **Wall thickness / mm** | 60.4 | 3.19 |
| **Internal diameter / m** | 1.94 | 0.43 |
| **Length / m** | NA | 47.24 |
| **Volume / $m^3$** | 357.50 | 64.76 |

Figures 1 and 2 depict 2-D section drawings of the new liquefaction and saccharification reactors, respectively.

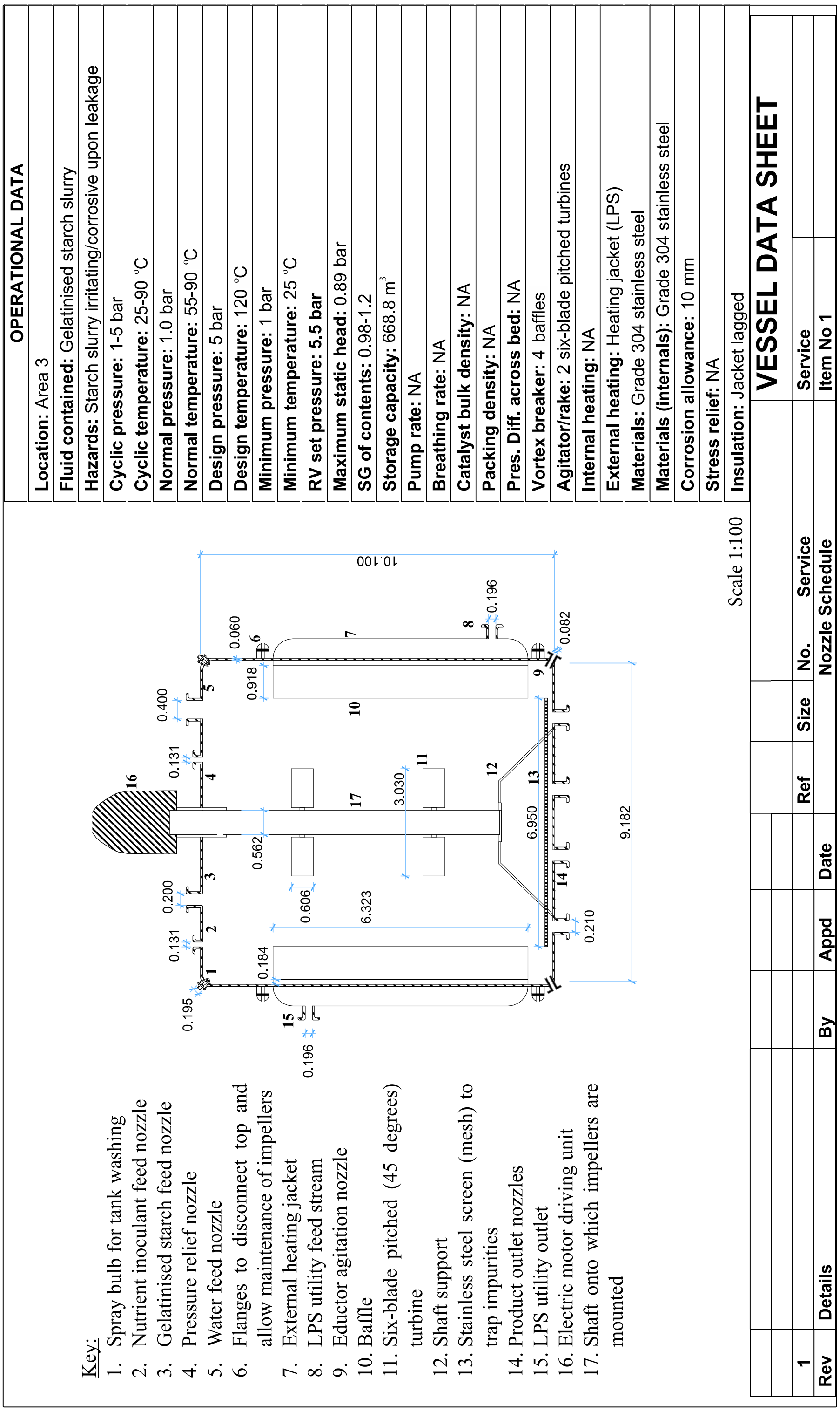



*Figure 1: 2-D section drawing of liquefaction reactor*

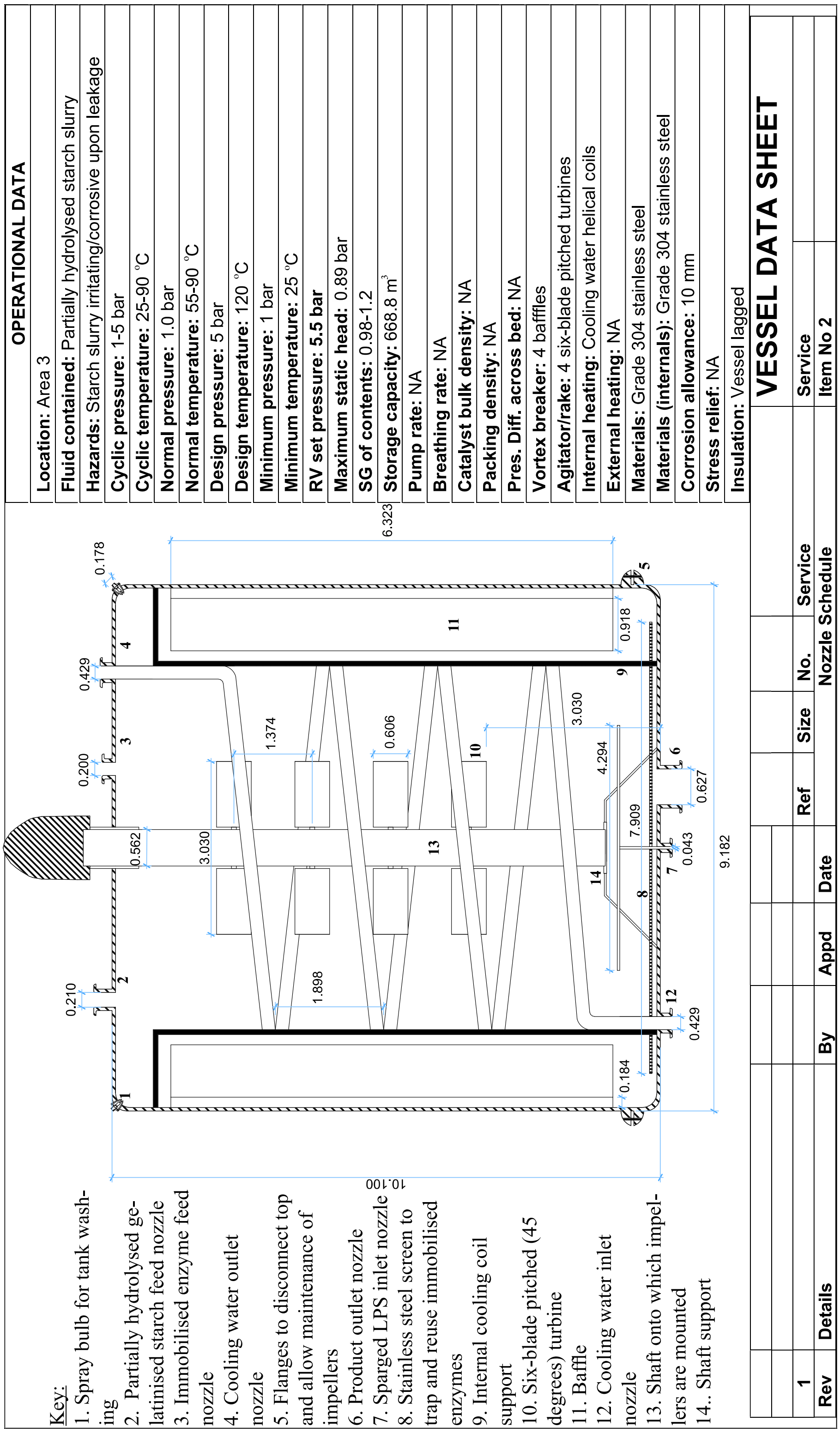



*Figure 2: 2-D section drawing of saccharification reactor*

The rotational speed of each of the impellers employed in each type of reactor was estimated at 66.3 rpm using the same assumptions employed in task B. In addition, the power requirement of the agitation system of each liquefaction and saccharification reactor was estimated at 0.88 and 1.77 MW, respectively, also using the same assumptions employed in task B. As a result, the overall agitation duties for the liquefaction reactors and saccharification reactors are approximately 3.52 MW and 28.32 MW. These carry a considerable degree of uncertainty since assumptions were made, which include a motor efficiency of 0.1 and minimal mass transfer limitation effects.

The agitation system duties will be provided by 3 phase power at 33 kV with a load capacity of 10 MVA.

Moreover, table 3 lists the estimated heating and cooling duty requirements of the 4 liquefaction and 16 saccharification reactors, respectively.

*Table 3: Heating and cooling duties estimated for the overall reactor system*

| | **4 Liquefaction reactors** | **16 Saccharification reactors** |
|---|---|---|
| **Duty / MW** | 6.5 | 14.5 |

The heating duty of each liquefaction reactor was estimated by assuming that the heat transfer coefficient of the LPS utility being carried in the heating jacket is much larger than that of the contents of the reactor. As a result, the overall heat transfer coefficient U was estimated at 254.75 W $m^{-2}$ $K^{-1}$ by use of equation 1 (Perry et al.,2008 p.18-23) and by assuming that the heat transfer resistance due to conduction is 0.0604/16.2, where 0.0604 is the liquefaction reactor wall thickness in m and 16.2 W $m^{-1}$ $K^{-1}$ (ASM Aerospace Specification Metals Inc. AISI Type 304 SS) is the thermal conductivity of grade 304 stainless steel. Equation 1 is relevant for pitched blade turbines.

(1)

In equation 1, h was estimated at 5091.3 W $m^{-2}$ $K^{-1}$. $Re_i$ is the impeller Reynolds number of 5000 as justified in task B. In addition, Pr was taken as the Prandtl number of water at 90 $^oC$. $D_j$ was taken as the diameter of the liquefaction reactor.

The heating duty Q of 1.625 MW for a single liquefaction reactor was estimated by assuming negligible enthalpy of reaction for liquefaction of gelatinised wheat starch (Akerberg et al. 2000). In addition, equation 2 was employed in calculating the duty of 1.625 MW. $\Delta T$ was specified at 35 $^oC$ as required for the liquefaction process (Akerberg et al. 2000). In equation 2, A represents the area for heat transfer between the heating jacket and the liquefaction reactor contents which was specified at 182.4 m

(2)

The LPS heating utility temperature change during heat transfer was also specified at 15 $^oC$. This allowed calculation of the mass flowrate of LPS required for a single liquefaction reactor by use of equation 3. In equation 3, w is the mass flowrate of LPS, and $C_p$ is the specific heat capacity of LPS at 131 $^oC$ of 2160 kJ $Kg^{-1}$ $K^{-1}$.

(3)

The mass flowrate of LPS required for a single liquefaction reactor was therefore estimated at 50.1 kg $s^{-1}$. These calculations carry with them a substantial degree of uncertainty of more than 60 %.

The cooling duty of each saccharification reactor was estimated in a similar manner by assuming that the heat transfer coefficient of the cooling water utility being carried in the helical cooling coils is much larger than that of the contents of the reactor. The overall heat transfer coefficient U in this case was estimated at 401.97 W $m^{-2}$ $K^{-1}$ by use of equation 4 (Perry et al., 2008 p. 18-24) and by assuming that the resistance due to conduction is 0.00319/16.2, where 0.00319 is the helical cooling coil wall thickness in m and 16.2 W $m^{-1}$ $K^{-1}$ (ASM Aerospace Specification Metals Inc. AISI Type 304 SS) is the thermal conductivity of grade 304 stainless steel.

(4)

In equation 4, $D_o$ is the outside diameter of the cooling coils of 0.436 m, $Re_i$ is the impeller Reynolds number specified at 5000, Pr is the Prandtl number of water at 55 $^oC$ of 3.2, $L_p$ is the impeller diameter of 3.03 m, k is the thermal conductivity of water at 55 $^oC$ of 0.645 W $m^{-1}$ $K^{-1}$, $D_j$ is the saccharification reactor diameter of 9.18 m and $n_b$ is the number of baffle type coils which was estimated at 20 by taking into account the pitch formed by the helical cooling coils of 1.898 m as also shown in figure 2. As a result, h was estimated at 437.7 W $m^{-2}$ $K^{-1}$. The cooling duty for a single saccharification reactor was estimated at 0.90 MW by use of equation 2. In this case A represents the area for heat transfer between the helical cooling coils and saccharification reactor contents which was specified at 64.2 $m^2$.

The cooling water utility temperature change during heat transfer was also specified at 20 $^oC$. This allowed calculation of the mass flowrate of cooling water required for a single saccharification reactor by use of equation 3. In this scenario, w is the mass flowrate of cooling water and $C_p$ is the specific heat capacity of cooling water at 18 $^oC$ of 4200 kJ $Kg^{-1}$ $K^{-1}$. The mass flowrate required for a single saccharification reactor was therefore estimated at 10.1 kg $s^{-1}$. These calculations carry with them a substantial degree of uncertainty of more than 70 % in this case.

The operating costs per year for the 20 reactors in total are listed in table 4, along with the utility requirements and utility costs. The annual costs were estimated by assuming plant operation of 8000 hours.

| Utility | Cost | Annual utility requirement | Annual operating costs |
|---|---|---|---|
| **Cooling water at 18 $^oC$** | 0.07 £ $t^{-1}$ | 4.65 x $10^9$ kg $y^{-1}$ | 3.26 x $10^5$ £ $y^{-1}$ |
| **LPS at 131 $^oC$** | 10 £ $t^{-1}$ | 5.8 x $10^8$ kg $y^{-1}$ | 5.77 x $10^6$ £ $y^{-1}$ |
| **Electricity from power grid** | 0.11 £ $(KWh)^{-1}$ | 254.72 GWh $y^{-1}$ | 28.02 x $10^6$ £ $y^{-1}$ |

*Table 4: Total operating costs for 20 reactors per annum*

Regarding the electricity supply from the grid, 3-phase power with a load capacity of 10 MVA will be available on site. The current capacity for the 3-phase power was calculated at 174.95 A assuming a typical power factor of 0.8 (http://www.mechreps.com/PDF/MRI_Formulas_Conversions.pdf).

Capital costs of the reactors and their internals were estimated by using the density of grade 304 stainless steel of 8060 kg $m^{-3}$(ASM Aerospace Specification Metals Inc. AISI Type 304 SS) and its cost of 2020.69 £ $t^{-1}$. Its price in USD of 2492 \$ $t^{-1}$ was acquired (https://worldsteelprices.com) which dates to 2019. This price was inflated to 2020 using a cost index ratio of 131.3/132.8 and converted to pounds using an exchange rate of £0.82 per USD.

Table 5 lists the estimated mass and capital costs of the liquefaction reactor internals including the reactors themselves. In estimating the mass of each component, simple geometric considerations were made. For example, the reactor was considered as an enclosed cylindrical surface of relevant thickness. This yields a substantial uncertainty equivalent to that of the relevant geometric approximations performed.

*Table 5: Capital cost estimates of 4 liquefaction reactors including their internal unit operations*

| Component | Number of units | Mass / t | Capital cost / £ |
|---|---|---|---|
| **Reactor body** | 4 | 1599 | 3.23 million |
| **Shaft** | 4 | 72.72 | 0.147 million |
| **Six-blade pitched turbines** | 8 | 1.066 | 2.15 thousand |
| **Heating jacket** | 4 | 710.3 | 1.44 million |
| **Baffles** | 16 | 75.62 | 0.153 million |

| | | | |
|---|---|---|---|
| **Total** | 24 | 2458.7 | 4.97 million |

In addition, table 6 lists the capital cost estimates of the saccharification reactors and their internals. These estimates were performed in a similar manner to those for the liquefaction reactors.

*Table 6: Capital cost estimates of 16 saccharification reactors including their internal unit operations*

| Component | Number of units | Mass / t | Capital cost / £ |
|---|---|---|---|
| **Reactor body** | 16 | 6396 | 12.9 million |
| **Shaft** | 16 | 285.2 | 0.576 million |
| **Six-blade pitched turbines** | 64 | 8.524 | 17.2 thousand |
| **Helical cooling coil system** | 16 | 28.37 | 57.3 thousand |
| **Baffles** | 64 | 302.5 | 0.611 million |
| **Support structure** | 16 | 102.2 | 0.207 million |

| | | | |
|---|---|---|---|
| **Total** | 192 | 7123 | 14.4 million |

## 3.2 Heat exchangers

The duties of E-302 and E-306 for heating stream 3.2 and cooling stream 3.7 were estimated at 39.36 and 28.75 MW, respectively. For E-302, equation 4 was employed, where $w_{3.2}$ is the mass flowrate of stream 3.2 and $C_{p3.2}$ is the specific heat capacity of stream 3.2 which was taken as 4.21 kJ $kg^{-1}$ $K^{-1}$.

The mass flowrate $w_{3.2}$ was estimated at 959521.23 kg $h^{-1}$. Stream 3.2 is towns water which will be provided on site. This mass flowrate is necessary to dilute stream 3.1 coming from Area 2 and acquire the desired $S_o$ of 60 g $dm^{-3}$ within each of the saccharification reactors, as justified in task B.

(4)

For E-306, equation 5 was employed. In equation 5, $w_{3.7}$ is the mass flowrate of stream 3.7 of 1197840 kg $h^{-1}$ and $C_{p3.7}$ is the specific heat capacity of stream 3.7 which was assumed to be that of water at 55 $^{o}$C of 4.18 kJ $kg^{-1}$ $K^{-1}$. This assumption was made because stream 3.7 is dilute in glucose.

(5)

The estimated duties carry a degree of uncertainty equivalent to that of the error in mass flowrates of streams 3.2 and 3.7 and the errors in the relevant specific heat capacities.

A shell and tube heat exchanger design was selected for both E-302 and E-306, since this type of heat exchanger is simple to design and cheap to manufacture (Sinnott et al., 2005).

Data sheets for E-302 and E-306 can be found in section 5, appendix 5.4. The relevant changes made to the heat exchangers when compared to the data sheets presented in task B, are the shell side nozzles which have been reduced in size. This was done to increase Reynolds number as well as better match the pipeline internal diameters of 0.40 m and 0.21 with that of the nozzles on the shell side. The pipeline diameters of 0.40 m and 0.21 m were selected for streams 3.2 and 3.7, respectively. Equations 6 and 7 indicate the estimated Reynolds numbers of streams 3.2 and 3.7.

(6)

(7)

In equations 6 and 7, $u_{3.2}$ and $u_{3.7}$ were estimated using the properties of water at 20 $^{o}$C and 55 $^{o}$C, respectively and the relevant pipeline diameters. For $u_{3.7}$, a density of 1020 kg $m^{-3}$ assigned due to the 6 % glucose weight fraction of the solution being transported (L. Torgesen et al., 1950).

A floating head type shell and tube heat exchanger incorporating 1 shell pass and 4 tube passes was specified for both duties. The floating head type shell and tube heat exchanger allows for differential expansion between the shell and tubes and easier cleaning and servicing (TEMA standards 9$^{th}$ edition, section 6 p.16). For E-302, the LPS utility was allocated to the tube side. This was also done for the cooling water utility being supplied to E-306.

The overall heat transfer coefficient for this kind of heat exchanger lies between 250-450 W $m^{-2}$ $K^{-1}$ (Sinnott et al., 2005 p.639).

Equation 8 was employed in calculating the area $A_{E\text{-}302}$ available for heat transfer in E-302. In equation 8, $\Delta T_{m,E\text{-}302}$ is the corrected log mean temperature difference incorporating the temperature of the LPS utility inlet and outlet as well as those of streams 3.2 and 3.3.

(8)

An overall heat transfer coefficient of 440 W $m^{-2}$ $K^{-1}$ was estimated which was off by 35 W $m^{-2}$ $K^{-1}$ from the final iteration performed.

The same method was employed in sizing E-306. Equation 9 was used in estimating the area $A_{E\text{-}306}$ available for heat transfer in E-306. An overall heat transfer coefficient of 415 W $m^{-2}$ $K^{-1}$ was estimated in this case which was off by 50 W $m^{-2}$ $K^{-1}$ from the final iteration performed.

(8)

In addition, table 7 lists the dimensions associated with components of E-302 and E-306. These carry an overall uncertainty of more than of 60 %, since the outlet temperature for LPS was estimated at 102.5 $^{o}$C at the specified heat transfer area of 1129.4 $m^2$, in Unisim®. This temperature was necessary to avoid any condensate from forming. The same process was carried out for the cooling water utility available for E-306. Its outlet temperature was estimated at 42 $^{o}$C.

*Table 6: Parameters associated with E-302 and E-306*

| **Parameter** | **Unit operation** | |
|---|---|---|
| | **E-302** | **E-306** |
| **Shell diameter / m** | 4.0 | 5.0 |
| **Shell wall thickness / mm** | 57.0 | 40.8 |
| **Tube length / m** | 8.0 | 10.9 |
| **Number of tubes** | 500 | 960 |
| **Internal tube diameter / m** | 0.088 | 0.115 |
| **External tube diameter / m** | 0.094 | 0.120 |

The shell thicknesses of each of E-302 and E-306 were estimated by assuming the equation for hoop stress in cylindrical pressure vessels and by using 80 % of the yield stress of high carbon steel of 224.8 MPa (ASM Aerospace Specification Metals Inc. AISI Type CS). This calculation was performed at the relevant design pressure of the shell components of E-302 and E-306.

The utility requirement for E-302 was estimated at 778 kg $s^{-1}$ of LPS provided at 131 $^{o}$C and 1.77 bar. In addition, that for E-306 was estimated at 277.8 kg $s^{-1}$ of cooling water provided at 18 $^{o}$C and 5 bar. These requirements were estimated in Unisim® by using the estimated parameter dimensions and duties of each heat exchanger.

Moreover, table 7 lists the total operating costs per annum associated with both heat exchangers along with the relevant utility costs. These were also estimated by assuming that the plant will be operational for 8000 h per year.

| **Utility** | **Cost / £ $t^{-1}$** | **Annual utility requirement / kg $y^{-1}$** | **Annual operating costs / £ $y^{-1}$** |
|---|---|---|---|
| **Cooling water at 18 °C** | 0.07 | $8.0 \times 10^{9}$ | $5.60 \times 10^{5}$ |
| **LPS at 131 °C** | 10 | $2.2 \times 10^{10}$ | $22.41 \times 10^{7}$ |

*Table 7: Combined operating costs per annum for E-302 and E-306*

The annual operating cost of E-302 was estimated at £ 222.41 million due to the fact that, a huge amount of towns water is being added to the liquefaction reactors so as to achieve the desired inlet concentration $S_o$ of 60 g $dm^{-3}$, which was justified in task B

The capital costs of E-302 and E-306 were estimated based on the area available for heat transfer, using formulae from the literature dating back to 2006 *(Seider et al.,2009 p.571-572).*

For E-302, equation 9 was employed, in which $C_P$ is the purchase cost of the heat exchanger. In equation 9, $F_M$ is unity due to the fact that, E-302 is made entirely out of high carbon steel (Seider et al.,2009 p.571). In addition, $F_L$ is unity because E-302 has tubes with associated length of greater than 20 ft (Seider et al.,2009 p.571 table 22-25).

$F_P$ which is a pressure correction factor, was calculated based on shell side pressure. This amounts to 5 bar or 72.51 psig for E-302. In equation 10, P is in psig.

Moreover, $C_B$ was calculated using equation 11. In equation 11, A is in $ft^2$. The heat transfer area associated with E-302 of 1129.4 $m^2$ was converted to 12 155 $ft^2$ for use in equation 11. In addition, equation 11 is applicable for the floating head type shell and tube heat exchanger (Seider et al.,2009 p.571 eqn. 22-39).

(9)

0.994 (10)

(11)

Using equations 9, 10 and 11, the purchase cost $C_P$ of E-302 was estimated at $ 92 582. This was converted to a 2020 cost by using the ratio of the cost indices for shell and tube heat exchangers between 2020 and 2006 of 532 (www.osti.gov/servlets/purl/797810/) and 500, respectively.

In addition, the exchange rate of £0.82 per USD (www.poundsterlinglive.com/best-exchange-rates/best-us-dollar-to-british-pound-history) was employed in order to estimate the capital cost of E-302. The relevant exchange rate was obtained on the 27 / 05 / 2020. As a result, the capital cost of E-302 was estimated at £ 81 263 using equation 12.

(12)

The exact same procedure was conducted for E-306 by converting the relevant heat transfer area of 874.6 $m^2$ to 9414.1 $ft^2$. In addition, the associated shell side inlet pressure of 1.6 bar was converted to 8.70 psig. The $F_P$ in this case was calculated at 0.98 and the base cost $C_E$ was calculated at \$ 71 500. This base cost was converted to £ 62 382 using the exact same method. Equation 13 was employed in calculating $CAPEX_{E\text{-}306}$ at £ 62 382.

(13)

The capital cost estimates of both E-302 and E-306 carry with them a substantial degree of uncertainty because correlations dating back to 2006 were involved, including the use of a cost indices which are more than 10 years apart.

## 3.3 Pumps

An additional set of centrifugal pumps has been incorporated in the new reactor model. This is E-303, which will pump the product of the liquefaction reactors to the saccharification reactors. Data sheets for both E-303 and E-305 sets of pumps can be found in section 5, appendix 5.5. The data sheet produced in task B for E-305 has changed. For the new E-305 data sheet, the assumption that the centrifugal pump will be pumping the product of saccharification to a storage vessel that is approximately 11 m high has been employed. This assumption considers the difference in batch times between Areas 3 and 4 of 2.5 h.

In estimating the new pump specifications, the assumption that the liquid level in each of the reactors is 9.18 m above ground level was employed. This incorporates the working volume of each of the reactors of 608 $m^3$.

Regarding the distance between the liquefaction and saccharification reactors, a pipeline length of 20 m was assumed. In addition, the pipelines were assumed to be made of carbon steel which has a roughness $\varepsilon$ of 0.05 mm (ASM Aerospace Specification Metals Inc. AISI Type CS). The internal diameter of the pipelines connecting liquefaction and saccharification reactors was specified at 0.21 m. The volumetric flowrate between liquefaction and saccharification was estimated at 1216.8 $m^3\ h^{-1}$ by dividing the overall mass flowrate resulting from all liquefaction reactors of 1197840 kg $h^{-1}$ by an assumed fluid density of 985 kg $m^{-3}$, as stated in task B. As a result, the estimated fluid velocity in the pipelines connecting liquefaction reactors to saccharification reactors was estimated at 9.76 m $s^{-1}$ by use of equation 14.

(14)

The Reynolds number of the flow $Re_{stream\ 3.4}$ which was assumed to be the same as that of stream 3.5, was estimated by use of equation 15.

(15)

In equation 15, the physical properties of water at 90 $^oC$ were assumed. The Darcy friction factor $f_{3.4}$ at the relevant $Re_{3.4}$ was estimated at 0.014. The Darcy friction factor was extracted from a moody diagram at a pipe roughness ratio $\varepsilon/D$ of 0.00024 (J. McGovern, 2011 p.2 figure 1). The velocity head lost due to friction $h_{f,3.4}$ along the pipeline connecting each liquefaction reactor to a saccharification reactor was estimated at 6.47 m by use of equation 16.

(16)

In determining the total dynamic head $H_{3.4}$ of the system, the total static head was added to the dynamic head (Perry et al. 2008 p. 10-22). The dynamic head was assumed to be the overall velocity head lost due to friction $h_{f,3.4}$. This can be assumed before installation of a pump in a system (Perry et al. 2008 p. 10-22 eqns. 10-41 & 10-44). The total static head was estimated by subtracting the differences in elevations between discharge and suction sides of the pump as well as adding an additional pressure head of 0.3 bar or 3.1 m. This was done to ensure that the pump does not cavitate. The total dynamic head $H_{3.4}$ of the pump was therefore estimated by adding the total static head and $h_{f,3.4}$ as shown in equation 17.

(17)

The net positive suction head available in this case $NPSH_{A,3.4}$ was estimated by subtracting the vapour pressure head $h_{v,3.4}$ which was estimated at the temperature of 90 $^oC$. This was calculated at 7.76 m by use of equation 18 (Perry et al., 2008 p.10-23 eqn. 10-55). In equation 18 (Perry et al.,2008 p.10-23 eqn.10-54), $p_{v,3.4}$ is the vapour pressure of water at 90 $^oC$ of 0.704 bar and $\rho_{3.4}$ is 985 kg $m^{-3}$, as aforementioned.

(18)

The power input $P_{3.4}$ requirement of E-303, was estimated by assuming a typical pump efficiency $\eta_{3.4}$ of 0.75. As a result, $P_{3.4}$ was estimated by use of equation 19 (Perry et al. 2008 p. 10-23 eqn. 10-53), where $Q_{3.4}$ is the volumetric flow rate of 1216.8 $m^3$ $h^{-1}$.

(19)

At the specified $Q_{3.4}$ of 1216.8 $m^3$ $h^{-1}$ and $H_{3.4}$ of 9.57 m, a single stage double suction centrifugal pump was identified with an impeller diameter of 0.85 m and a rotational speed of 1200 rpm (www.shinkohir.co.jp/pdf/catalog/Centrifugal Pumps, p.13, GHD-400 model).

These estimates carry with them a relatively large degree of uncertainty because $NPSH_{A,3.4}$ was not calculated in the scenario that a pump installation is being designed. In addition, $f_{3.4}$ was estimated from a chart.

The same analysis was conducted for the set of centrifugal pumps E-305. The relevant Darcy friction factor was also estimated at approximately 0.014 at an $Re_{3.6}$ of 5933536 by assuming the physical properties of water at 55 $^oC$. In this scenario though, the resultant glucose solution will be pumped from a height of 9.18 m within the saccharification reactors to a storage vessel at a new liquid surface level of 11 m. In addition, the glucose solution will flow through a heat exchanger E-306 through which a pressure drop of 0.6 bar was specified.

The pipe diameter of stream 3.6 was also assumed to be 0.21 m and the pipeline length between the saccharification reactors and storage vessel was assumed to be 30. The required volumetric flowrate was also assumed to be 1216.8 $m^3$ $h^{-1}$ despite the temperature difference between streams 3.6 and 3.4. The total velocity head lost due to friction in this case $h_{f,3.6}$ was estimated at 12.94 m. In addition, the total dynamic head $H_{3.6}$ was estimated at 19.08 m and the relevant net positive suction head $NPSH_{A.3.6}$ was estimated at 18.96 m. The vapour pressure employed in this calculation $p_{v,3.6}$ was taken as that of water at 55 °C of 0.12 bar. As a result, the power input requirement $P_{3.6}$ of E-305 was estimated at 83.09 kW by use of equation 20 and a typical pump efficiency $\eta_{3.6}$ of 0.75.

(20)

The same single stage double suction pump design as that for E-303 was identified but with a different power requirement of 83.09 kW.

The required duty of both E-303 and E-305 pump sets will also be provided by 3 phase power at 33kV with a load capacity of 10 MVA. Table 8 lists the associated operating costs for each set of centrifugal pumps. These were estimated by assuming 8000 h of plant operation per annum as well as an electricity cost of 0.11 £ $(KWh)^{-1}$

| **Pump set** | **Annual utility requirement / GWh $y^{-1}$** | **Annual operating costs / £ $y^{-1}$** |
|---|---|---|
| **E-303** | 0.3328 | 36 608 |
| **E-305** | 0.6647 | 73 117 |

*Table 8: Operating costs for E-303 and E-305 per annum*

In estimating the capital costs of each set of centrifugal pumps, cost correlations were employed. These include a size factor S which is dependent on flowrate Q and pump head H, a base cost $C_B$ dating back to 2006 and material and pump-type correction factors $F_M$ and $F_T$, respectively (Seider et al.,2009 p.560-561). Equations 21, 22 and 23 were employed to estimate the purchase cost of both E-303 and E-305 dating back to 2006.

(21)

(22)

(23)

In equation 21, Q is in gallons per minute and H is in feet. For E-303, $S_{E\text{-}303}$ was estimated at 30 020 gpm $ft^{0.5}$ by converting 1216 $m^3$ $h^{-1}$ to 5357.4 gpm and H of 9.57 m to 39.40 ft. As a result, $C_{B,\ E\text{-}303}$ was calculated at \$ 8388.1 by use of equation 22. $F_{t,\ E\text{-}303}$ was taken as 2 (Seider et al.,2009 p.561 table 22-20) and $F_{M,E\text{-}303}$ was taken as 1 (Seider et al.,2009 p.562 table 22-21). $C_{p,\ E\text{-}303}$ was therefore estimated at \$ 16 334.8. This was converted to a 2020 purchase cost by using the ratio of the cost indices for centrifugal pumps between 2020 and 2006 of 532 and 500, respectively. The relevant purchase cost is \$ 17 380.3. The exchange rate of £0.82 per USD was also employed to estimate $CAPEX_{E\text{-}303}$ at £ 14 251.8

The same calculations were performed for E-305. In this case, the associated Q is also 5357.4 gpm but the associated H is 19.08 m which was converted to 62.60 ft for use in equation 21. As a result, $S_{E\text{-}305}$ was estimated at 42 388 gpm $ft^{0.5}$. $C_{B,\ E\text{-}305}$ was calculated at \$ 9855.8, using equation 22. $F_{t,\ E\text{-}305}$ was also taken as 2 (Seider et al.,2009 p.561 table 22-20) and $F_{M\text{-}E\text{-}305}$ was taken as 1 (Seider et al.,2009 p.562 table 22-21). $C_{p,E\text{-}305}$ was therefore calculated at \$ 19 711.7. $CAPEX_{E\text{-}305}$ was eventually estimated at £ 16 163.6 by using the same exchange rate of £0.82 per USD.

## 3.4 Equipment lists for Area 3

*Table 8: Heat exchanger equipment list*

| Tag number | Description | No. of units | Total duty (MW) | No. of cold streams | No. of hot streams | Duty per stream (MW) | Area per stream ($m^2$) | Cold stream ΔP (bar) | Cold stream ΔT (ºC) | Hot stream ΔP (bar) | Hot stream ΔT (ºC) | Minimum temperature approach (ºC) | Hot stream location (shell/tube) | Cold stream location (shell/tube) | Shell material | Tube material | Notes | Calculation reference | Tota area $m^2$ |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| 3 | Shell and tube heat exchanger type AES | 1 | 39.36 | 1,0 | 1 | 39.36 | 1129.4 | 0.70 | 35,0 | 0.77 | 28.5 | 47.5 | tube | shell | carbon steel | carbon steel | Long tubes, 8 m | Section 3, part 3.2 | 1129. |
| 5 | Shell and tube heat exchanger type AES | 1 | 28.75 | 1,0 | 1 | 28.75 | 1129.4 | 0.50 | 24,0 | 0.60 | 20.0 | 13,000 | shell | tube | carbon steel | carbon steel | Long tubes, 10.9 m | Section 3, part 3.2 | 874.6 |

*Table 9: Pump equipment list*

| Tag number | Description | No. of units | Type | Volumetric flowrate ($m^3 h^{-1}$) | Mass flowrate ($kg h^{-1}$) | ΔP (bar) | ΔT (ºC) | Head (m) | Materials | Driver | Notes | Calculation reference | Driver power (kW) |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| 6 | Single stage double suction | 4 | Centrifugal | 1216.8 | 1198548 | 0.93 | 0.00 | 9.58 | Carbon steel | Electric motor | Estimated rpm of 1200 $min^{-1}$ | Section 3, part 3.3 | 41.6 |
| 4 | Single stage double suction | 4 | Centrifugal | 1216.8 | 1198548 | 0.60 | 0.00 | 19.08 | Carbon steel | Electric motor | Estimated rpm of 1200 $min^{-1}$ | Section 3, part 3.3 | 83.0 |

*Table 10: Vessel equipment list*

| Tag number | Description | No. of units | Diameter (m) | Total vessel length (m) | Operating temperature (ºC) | Total pressure drop (bar) | Details of internals | Materials | Calculation reference | Weight (tonnes) |
|---|---|---|---|---|---|---|---|---|---|---|
| 1 | Stirred tank vessel with external heating jacket. | 4 | 9.18 | 10.1 | 90 | NA | 4 baffles, 2 six-blade pitched turbines | Grade 304 stainless steel | Section 3, part 3.1 | 614.75 |
| 2 | Stirred tank vessel with internal helical cooling coils | 16 | 9.18 | 10.1 | 55 | NA | 4 baffles, 4 six-blade pitched turbines | Grade 304 stainless steel | Section 3, part 3.1 | 445.19 |

# 4 Additional relevant information associated with Area 3 for further analysis

## 4.1 Area 3 layout sketches

A plan view and side elevation view sketch of Area 3 can be found in section 5, appendix 5.6

## 4.2 Summary of utility requirements in Area 3

Table 11 lists the utility requirements for the process equipment located in Area 3. The utility requirements were estimated by assuming 8000 h of plant operation per year. In table 11, the electricity requirement per unit operation in KWh $y^{-1}$ incorporates the 8000 h of annual plant operation. In addition, E-301 considers all 4 liquefaction reactors and E-304 accounts for all 16 saccharification reactors.

| Unit operation | Utility requirement per annum | | |
|---|---|---|---|
| | **LPS at 131 °C / t $y^{-1}$** | **Cooling water at 18 °C / t $y^{-1}$** | **Electricity from grid KWh $y^{-1}$** |
| **E-301** | 580000 | NA | 28160000 |
| **E-302** | 22000000 | NA | NA |
| **E-303** | NA | NA | 332800 |
| **E-304** | NA | 4650000 | 226560000 |
| **E-305** | NA | NA | 664000 |
| **E-306** | NA | 8000000 | NA |

*Table 11: Annual utility requirements for Area 3-unit operations*

In Area 3, towns water provided at 20 °C is also required. This is necessary to dilute the stream entering from Area 2 to achieve the desired saccharification reactor inlet $S_o$ of 60 g $dm^{-3}$. As a result, towns water is an additional utility requirement for Area 3.

The mass requirement of towns water was estimated at 9595.2 t for every 10 h of batch cycle time, as also shown on the PFD in section 5, appendix 5.1. An equivalent towns water mass flow per year of 7676160 t $y^{-1}$ is therefore required in Area 3.

## 4.3 Summary of process and material hazards in Area 3

The hazardous materials associated with Area 3 include the LPS utility being supplied at 131 °C and 1.77 bar and the hydrolysed starch slurry that exits the liquefaction reactors. The NaOH $_{(aq)}$ and $H_2SO_4$ $_{(aq)}$ solutions being supplied to the saccharification reactors are also considered hazardous.

| Material | Associated hazards and their potential consequences |
|---|---|

| | |
|---|---|
| Hydrolysed starch slurry | In the scenario of loss of containment from pipes or reactors, it can cause skin and eye irritation upon physical contact. During downtime due to maintenance of the reactors, workers are exposed to this danger. In addition, the associated material can cause respiratory breathing problems |
| Low pressure steam | In the case of leakage, LPS can cause severe burns and cause pipes to become abnormally warm as well as increase the local ambient temperature. This can also cause loss of consciousness for workers nearby |
| NaOH (aq) | Highly corrosive solution of caustic soda can cause severe skin burns upon physical contact following leakage. In addition, it may corrode adjacent pipework or vessel walls leading to unwanted loss of containment of another material |
| $H_2SO_4$ (aq) | Highly corrosive solution. Can cause skin burns and eye damage in the case of physical contact in the workplace following loss of containment. In addition, it can corrode adjacent pipework. Moreover, the substance is harmful to aquatic life and as a result, effluent treatment is necessary prior to discharge to the environment |

*Table 12: Area 3 material hazards*

The main process hazards associated with Area 3 include large volumetric flowrates in certain pipelines and large utility mass flowrates in certain pieces of process equipment such as heat exchangers E-302 and E-306. Table 13 summarises the process hazards relevant to Area 3.

*Table 13: Area 3 process hazards*

| **Process** | **Associated hazards and their potential consequences** |
|---|---|
| Streams 3.4, 3.5, 3.6, 3.7 | The diameter of the pipelines carrying these streams is 210 mm. The fluid velocity is high, and the solution being transferred fouls significantly. Blockage of these pipelines due to fouling or accumulation of solids can cause rupture and loss of containment. In addition, streams 3.4 and 3.5 are at a temperature of 90 $^oC$. The associated pipelines are therefore very warm, and this may put nearby workers at risk of suffering from second degree burns |
| E-302 | The mass flowrate of LPS on the tube side is significantly large at 778 kg $s^{-1}$. In addition, E-302 consists of 8 m long tubes. A relief, throttling valve is necessary on the tube side of E-302. The pressure on the shell side is moderately large at 5 bar. In case of rupture in the tube side, there is risk of instant release of LPS and of E-302 collapsing |

| | |
|---|---|
| E-301 | The liquefaction reactors are under semi-batch operation. They will be filled sequentially by diverting the incoming streams 3.2 and 3.1 every 30 minutes by use of the programmable logic controller. In case the controller fails, there is risk of the liquefaction reactors overfilling and causing flooding. The resulting solution at 90 $^{o}$C poses a threat to the surrounding equipment and workforce. |

## 4.4 Summary of relief stream data in Area 3

A drain system is in place in Area 3 which removes the untreated aqueous effluent from the pipework and all 20 reactors. This system, for which a section of the relevant pipeline is shown on the saccharification reactor P&ID in section 5, appendix 5.3, transports aqueous effluent containing unconverted oligosaccharides and other impurities such as proteins, to a waste treatment facility. The untreated aqueous effluent is decontaminated, prior to discharge to the environment, and the relevant treatment is carried out by a different company at a cost of £ 1.90 per tonne of aqueous effluent.

In addition, a relief system is in place which mainly deals with the gas exhaust from the saccharification reactors. The relevant gas exhaust consists of low-pressure steam which is used to sterilise the saccharification reactors prior to each batch. This relief system connects both liquefaction and saccharification reactors and leads to an exhaust chimney/flue on site. The exhaust flue is necessary to avoid sudden condensation of LPS upon exposure to ambient atmosphere. The exhaust gas is also sterilised prior to discharge by use of monofilament polypropylene filters of appropriate pore sizes (S. Ramaswamy et al., 2013). Moreover, vacuum relief is in place, and since, non-flammable fluids are being processed, ambient air is being used for this purpose.

The fibre being removed from the liquefaction reactors during cleaning is recycled and temporarily stored. It is then offered to paper-manufacturing companies. In case, this does not happen, it is disposed of in a nearby landfill site. This is considered reasonable since, it takes less than a year for it to degrade (U. Tchirner et al., 2007).

## 4.5 Analysis of heat integration opportunities in Area 3

Table 14 lists the streams associated with Area 3 and their relevant temperatures

*Table 14: Area 3 stream temperatures*

| Stream | Temperature / $^{o}$C |
|---|---|
| 3.1 | 55 |
| 3.2 | 20 |
| 3.3 | 55 |
| 3.4 | 90 |
| 3.5 | 90 |
| 3.6 | 55 |
| 3.7 | 55 |

| | |
|---|---|
| 3.8 | 35 |

As far as heat integration is concerned, a proper analysis using an energy cascade approach and a relevant grand composite plot, was not conducted. There was not enough time to do so, and the results of such an analysis would have been erroneous since, empirical correlations for heat capacities of relevant streams at their starting and target temperatures were not found in literature.

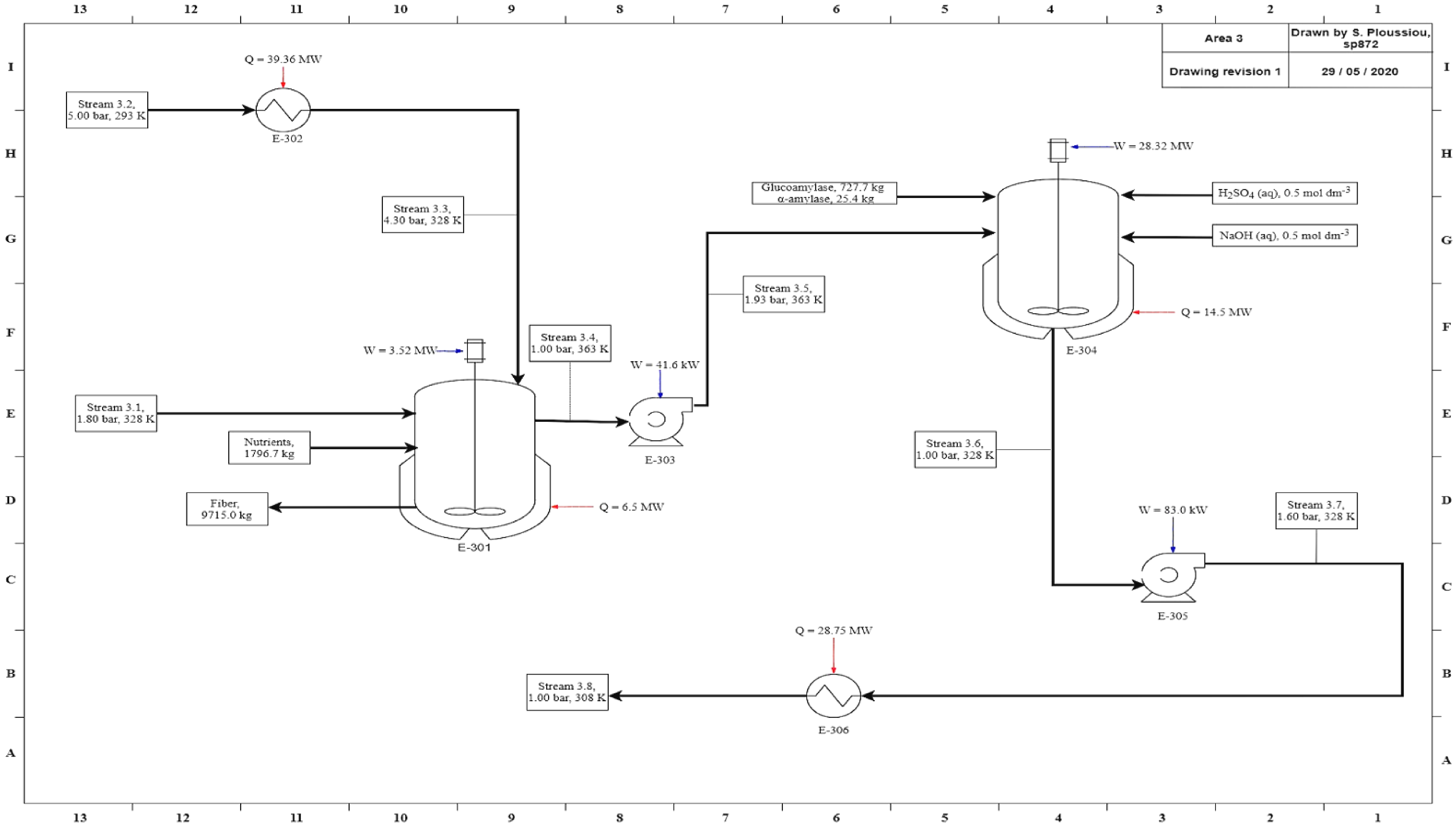
Nevertheless, there are opportunities in terms of heat integration in Area 3. Stream 3.2 consisting of towns water at 20 $^oC$ is being heated to 55 $^oC$ by 778 kg $s^{-1}$ of LPS provided at 131 $^oC$. There is scope to eliminate this massive utility requirement by heat integrating this stream with stream 3.5, which is the one exiting E-303.

A minimum temperature approach $\Delta T_{min}$ of 10 $^oC$ can be used in a heat integration scheme incorporating streams 3.2 and 3.5 since, liquids are involved. A heat integration scheme between these streams would reduce if not even eliminate the overall utility requirement needed to heat towns water to 55 $^oC$.

In addition, the proposed heat recovery scheme would also reduce the utility requirement for the saccharification reactors since, stream 3.5 would be cooled down to less than 90 $^oC$ in this scenario thus, eliminating some of the cooling duty requirement of the saccharification reactors.

# 5 Appendix

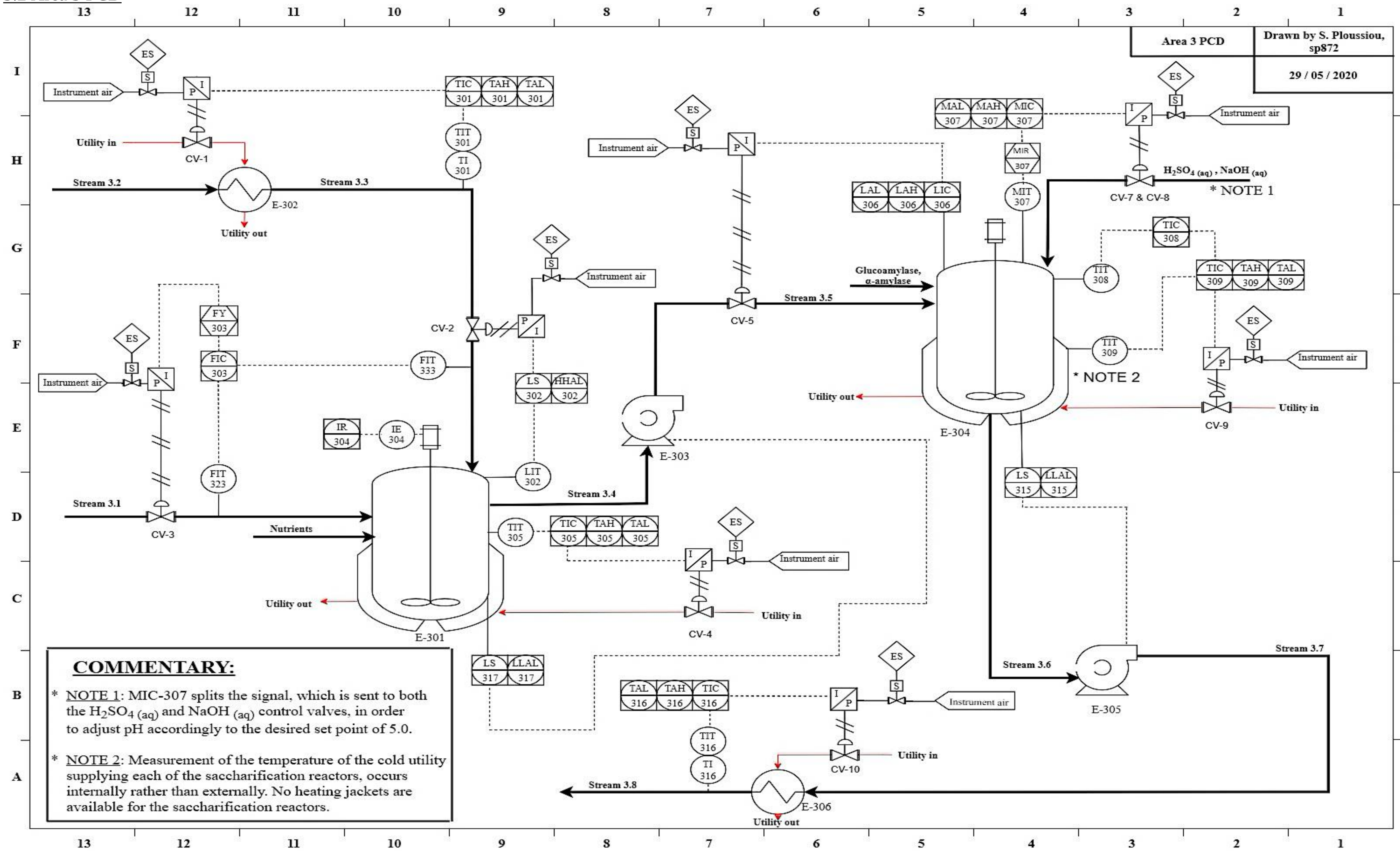
5.1 Area 3 PFD and stream tables



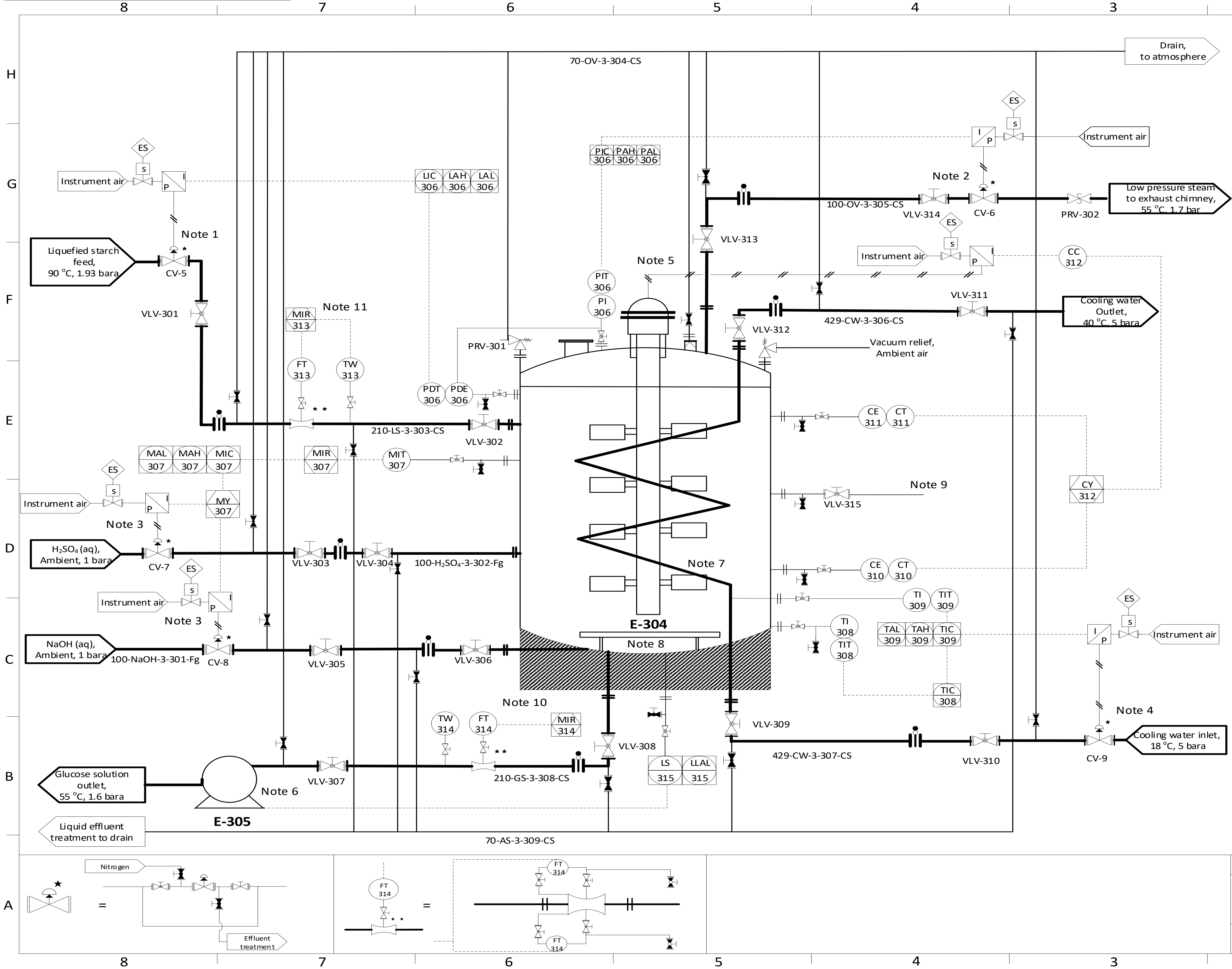
## AREA 3 STREAM TABLES

| Line number | 3.1 | 3.2 | 3.3 | 3.4 | 3.5 | 3.6 | 3.7 | 3.8 |
|---|---|---|---|---|---|---|---|---|
| **Mass Flow (t)** | 2383.2 | 9595.2 | 9595.2 | 11970.4 | 11970.4 | 11970.4 | 11970.4 | 11970.4 |
| **Batch cycle time (h)** | 10 | 10 | 10 | 10 | 10 | 10 | 10 | 10 |
| **Composition (wt%)** | | | | | | | | |
| **Starch (gelatinised)** | 0.30000 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| **Water** | 0.64130 | 1 | 1 | 0.92887 | 0.92887 | 0.92857 | 0.92857 | 0.92857 |
| **Trace impurities** | 0.05710 | 0 | 0 | 0.01140 | 0.01140 | 0.01140 | 0.01140 | 0.01140 |
| **α-amylase** | 0.00003 | 0 | 0 | 0.00003 | 0.00003 | 0.00003 | 0.00003 | 0.00003 |
| **Glucose** | 0.00160 | 0 | 0 | 0.00760 | 0.00760 | 0.06000 | 0.06000 | 0.06000 |
| **Maltose** | 0 | 0 | 0 | 0.00600 | 0.00600 | 0 | 0 | 0 |
| **Maltotriose** | 0 | 0 | 0 | 0.00030 | 0.00030 | 0 | 0 | 0 |
| **Maltotetraose** | 0 | 0 | 0 | 0.00110 | 0.00110 | 0 | 0 | 0 |
| **Maltopentaose** | 0 | 0 | 0 | 0.00040 | 0.00040 | 0 | 0 | 0 |
| **Maltohexaose** | 0 | 0 | 0 | 0.00200 | 0.00200 | 0 | 0 | 0 |
| **Maltoheptaose** | 0 | 0 | 0 | 0.00080 | 0.00080 | 0 | 0 | 0 |
| **Oligosaccharides (DP>7)** | 0 | 0 | 0 | 0.04150 | 0.04150 | 0 | 0 | 0 |
| **Phase** | Solid, Liquid | Liquid | Liquid | Liquid | Liquid | Liquid | Liquid | Liquid |
| **Pressure (bar)** | 1.80 | 5.00 | 4.30 | 1.00 | 1.93 | 1.00 | 1.60 | 1.00 |
| **Temperature (K)** | 328 | 293 | 328 | 363 | 363 | 328 | 328 | 308 |

## 5.2 Area 3 PCD


Area 3 PCD
Drawn by S. Ploussiou, sp872
29 / 05 / 2020
Instrument air
Utility in
Utility out
CV-1
E-302
Stream 3.2
Stream 3.3
TIC 301
TAH 301
TAL 301
TIT 301
TI 301
CV-2
FY 303
FIC 303
FIT 333
FIT 323
LS 302
HHAL 302
LIT 302
IR 304
IE 304
Stream 3.1
CV-3
Nutrients
Stream 3.4
E-301
TIT 305
TIC 305
TAH 305
TAL 305
CV-4
LS 317
LLAL 317
E-303
CV-5
Stream 3.5
LAL 306
LAH 306
LIC 306
MAL 307
MAH 307
MIC 307
MIR 307
MIT 307
$H_2SO_{4\ (aq)}$, $NaOH_{(aq)}$
* NOTE 1
CV-7 & CV-8
Glucoamylase, α-amylase
TIC 308
TIT 308
TIC 309
TAH 309
TAL 309
TIT 309
* NOTE 2
CV-9
E-304
LS 315
LLAL 315
Stream 3.6
Stream 3.7
E-305
TAL 316
TAH 316
TIC 316
TIT 316
TI 316
CV-10
Stream 3.8
E-306
ES
COMMENTARY:
* NOTE 1: MIC-307 splits the signal, which is sent to both the $H_2SO_{4\ (aq)}$ and $NaOH_{\ (aq)}$ control valves, in order to adjust pH accordingly to the desired set point of 5.0.
* NOTE 2: Measurement of the temperature of the cold utility supplying each of the saccharification reactors, occurs internally rather than externally. No heating jackets are available for the saccharification reactors.

## 5.3 Saccharification reactor P&ID




## Commentary

### Control

1 - Loop controls liquid level in E-303, by manipulating the liquified starch feed flowrate via CV-5. This is performed by PDT-306 sending a signal to LIC-306. PDE-306 measures pressure difference between liquid and overhead vapour space in this control loop.

2 - Loop controls pressure by manipulating low pressure steam flowrate in exhaust gas line leading to chimney via CV-6. This control loop and relevant stream exist in addition to other pressure relief devices in the system.

3 - Loop controls pH in E-303, by manipulating CV-7 and CV-8. MIR-307 linearises the pH measurement made by MIT-307 by taking the antilog of the measurement. The converted signal is then sent to MIC-307, the output of which is split. In the case that pH drops below the set point of 5, CV-8 is opened.

4 - Loop controls temperature of liquid phase in E-303 by manipulating cooling water flowrate via CV-9. Cascade control is employed in this case in order to minimise lag time between disturbances in cooling water flowrate and temperature correction in E-303.

5 - Loop controls the composition of glucose in E-303 by manipulating agitator speed. CE-310 and CE-311 measure glucose composition at two different positions in E-303. CY-312 compares the measurements and sends a signal to CC-312, the output of which acts to adjust agitator power by manipulating the current supplied to it.

### Trips

6 - Low level trip to E-304.
In addition, high and low pressure, level, temperature and pH alarms are shown on the P&ID, which in case of an emergency, would trigger operator action.
A FULL area 3 trip would automatically trigger an emergency shutdown.

### Emergency Shutdown Actions

CV-5 shuts
CV-6 opens fully
CV-7 & CV-8 shut
CV-9 shuts
E-304 no action. Continues unless low level trip in note 6 is active

### General

7 - Not actual layout of internal cooling coils in E-303. The helical cooling coil entrance to the reactor is on the opposite bottom end to the one shown on the P&ID.

8 - A sparger for LPS at 131 °C, is positioned at the bottom of the reactor. This is not shown on the P&ID since it's purpose is to steam and therefore sterilise all the ports and walls of the reactor prior to each charge of feed.

9 - Small pipe section connected to E-303, consisting of a drainage valve and manual isolation valve, VLV-315. It permits E-303 inventory samples to be extracted for laboratory analysis

10 - FT-314 and TW-314 measure the pressure difference and liquid temperature, respectively. These readings are fed to MIR-314 which analyses and converts them to flowrate (Glucose solution outlet flowrate).

11 - MIR-313 performs the same analysis and conversion as MIR-314 in note 10, but in this scenario the measured flowrate is that of the Liquefied starch feed.

**850 – OV – 2 – xxx – CS**
**Nominal diameter (mm) – Material Contained – Area – Unique line number – Pipe material**

**Material Key:**

**NaOH = Sodium hydroxide** **AS=Aqueous slurry**
**$H_2SO_4$ = Sulfuric acid** **GS = Glucose solution**
**CS = Carbon Steel** **Fg = Fiberglass**

Area 3- Saccharification reactor

**Drawn by:**
S. Ploussiou, sp872

Revision: 1 Date: 29th May 2020

## 5.4 E-302 and E-306 Data Sheets

### E-302 HEAT EXCHANGER DATA SHEET

| | | |
|---|---|---|
| **Client** CEB | **Area 3** | **Item no. 3** |

| | | |
|---|---|---|
| **Service** Towns water heating from 20 C to 55 C | | |
| **No of units 1** | **Type** AES | **Orientation** Horizontal |

| | Units | Shell Fluid (Towns water) | | Tube Fluid (LPS) | |
|---|---|---|---|---|---|
| | | **In** | **Out** | **In** | **Out** |
| **Total fluid flow** | **kg $s^{-1}$** | 266.5 | 266.5 | 778 | 778 |
| **Liquid** | **kg $s^{-1}$** | 266.5 | 266.5 | 0 | 0 |
| **Vapour** | **kg $s^{-1}$** | 0 | 0 | 778 | 778 |
| **Non-condensable** | **kg $s^{-1}$** | 0 | 0 | 0 | 0 |
| **Steam** | **kg $s^{-1}$** | 0 | 0 | 778 | 778 |
| **Water** | **kg $s^{-1}$** | 266.5 | 266.5 | 0 | 0 |
| **Temperature** | **C** | 20.0 | 55.0 | 131.0 | 102.5 |
| **Mol. Wt of vapour + NC** | | 0 | 0 | 1.0 | 1.0 |
| **Liquid density** | **kg $m^{-3}$** | 998 | 985 | 0 | 0 |
| **Liquid viscosity** | **Pa s** | 0.0010 | 0.0005 | 0 | 0 |
| **Vapour + NC viscosity** | **Pa s** | NA | NA | 0.000013 | 0.000012 |
| **Liquid sp. Heat** | **kJ $kg^{-1}$ $K^{-1}$** | 4.18 | 4.21 | NA | NA |
| **Vapour + NC sp. Heat** | **kJ $kg^{-1}$ $K^{-1}$** | NA | NA | 2.16 | 2.10 |
| **Liquid thermal conductivity** | **W $m^{-1}$ $K^{-1}$** | 0.603 | 0.649 | NA | NA |
| **Vap. + NC thermal conduct.** | **W $m^{-1}$ $K^{-1}$** | NA | NA | 0.0274 | 0.0245 |
| **Latent heat** | **kJ $kg^{-1}$** | 2453.5 | 2282.5 | 2211 | 2257 |
| **Surface tension** | **N $m^{-1}$** | 0.0723 | 0.0669 | NA | NA |
| **Dew pt/bubble pt** | **C** | 151.8 | 145.0 | 115.0 | 100.0 |
| **Freeze pt/pour pt** | **C** | 0 | 0 | 0 | 0 |
| **Inlet pressure** | **bar** | 5 | | 1.77 | |
| **Allowable press. Drop** | **bar** | 0.70 | | 0.77 | |
| **Fouling resistance** | **$m^2$ K $W^{-1}$** | 0.00025 | | 0.00009 | |
| **Heat exchanged** | **MW** | 39.36 | | | |
| **Min./max. operating temp** | **C** | 15 | 60 | 95 | 135 |
| **Max. operating pressure** | **bar** | 6 | | 3 | |
| **Min. design pressure** | **bar** | 5 | | 2 | |
| **Relief valve set pressure** | **bar** | 6 | | 3 | |
| **Design pressure** | **bar** | 7 | | 4 | |
| **Cyclic service** | | ☒ | | ☒ | |
| **Materials** | | Carbon steel | | Carbon steel | |
| **Corrosion allowance** | **mm** | 1.0 | | 1.0 | |
| **Line size** | **mm** | 400 | 400 | 545 | 545 |
| **Heat treatment** | | ☒ | | ☒ | |
| **Hazards** | | Moderately pressurised | | Hot upon physical contact | |
| **Insulation** | | NA | | NA | |
| **Cleaning** | | Once every 3 months | | Once every 3 months | |

## E-306 HEAT EXCHANGER DATA SHEET

| | | |
|---|---|---|
| **Client** CEB | **Area** 3 | **Item no.** 5 |

| | | |
|---|---|---|
| **Service** Glucose solution cooldown from 55 C to 35 C | | |
| **No of units 1** | **Type** AES | **Orientation** Horizontal |

| | **Units** | **Shell Fluid (Glucose solution)** | | **Tube Fluid (Cooling water)** | |
|---|---|---|---|---|---|
| | | **In** | **Out** | **In** | **Out** |
| **Total fluid flow** | **kg $s^{-1}$** | 332.7 | 332.7 | 277.8 | 277.8 |
| **Liquid** | **kg $s^{-1}$** | 332.7 | 332.7 | 277.8 | 277.8 |
| **Vapour** | **kg $s^{-1}$** | 0 | 0 | 0 | 0 |
| **Non-condensable** | **kg $s^{-1}$** | 0 | 0 | 0 | 0 |
| **Steam** | **kg $s^{-1}$** | 0 | 0 | 0 | 0 |
| **Water** | **kg $s^{-1}$** | 310.0 | 310.0 | 277.8 | 277.8 |
| **Temperature** | **C** | 55 | 35 | 18 | 42 |
| **Mol. Wt of vapour + NC** | | NA | NA | NA | NA |
| **Liquid density** | **kg $m^{-3}$** | 1020.0 | 1022.0 | 999.0 | 992.0 |
| **Liquid viscosity** | **Pa s** | 0.00050 | 0.00072 | 0.00105 | 0.00063 |
| **Vapour + NC viscosity** | **Pa s** | NA | NA | NA | NA |
| **Liquid sp. Heat** | **kJ $kg^{-1}$ $K^{-1}$** | 4.182 | 4.180 | 4.182 | 4.180 |
| **Vapour + NC sp. Heat** | **kJ $kg^{-1}$ $K^{-1}$** | NA | NA | NA | NA |
| **Liquid thermal conductivity** | **W $m^{-1}$ $K^{-1}$** | 0.6485 | 0.6251 | 0.6002 | 0.6340 |
| **Vap. + NC thermal conduct.** | **W $m^{-1}$ $K^{-1}$** | NA | NA | NA | NA |
| **Latent heat** | **kJ $kg^{-1}$** | 2241 | 2277 | 2130 | 2142 |
| **Surface tension** | **N $m^{-1}$** | 0.06685 | 0.07036 | 0.07331 | 0.06914 |
| **Dew pt/bubble pt** | **C** | 109.0 | 98.0 | 151.8 | 147.9 |
| **Freeze pt/pour pt** | **C** | -5 | -5 | 0 | 0 |
| **Inlet pressure** | **bar** | 1.6 | | 5 | |
| **Allowable press. Drop** | **bar** | 0.6 | | 0.5 | |
| **Fouling resistance** | **$m^2$ K $W^{-1}$** | 0.0003 | | 0.0002 | |
| **Heat exchanged** | **MW** | 28.75 | | | |
| **Min./max. operating temp** | **C** | 30 | 60 | 13 | 47 |
| **Max. operating pressure** | **bar** | 3 | | 6 | |
| **Min. design pressure** | **bar** | 2 | | 5 | |
| **Relief valve set pressure** | **bar** | 3 | | 6 | |
| **Design pressure** | **bar** | 4 | | 7 | |
| **Cyclic service** | | ⊠ | | ⊠ | |
| **Materials** | | Carbon Steel | | Carbon Steel | |
| **Corrosion allowance** | **mm** | 1.1 | | 1.0 | |
| **Line size** | **mm** | 400 | 400 | 371 | 371 |
| **Heat treatment** | | ⊠ | | ⊠ | |
| **Hazards** | | Respiratory irritation | | Moderately pressurised | |
| **Cleaning** | | Once every 3 months | | Once every 3 months | |

## 5.5 E-303 and E-305 Data Sheets

### E-303 Centrifugal Pump Data Sheet

| | | |
|---|---|---|
| **Client** CEB | **Area** 3 | **Item no. 6** |

| **Service** | | |
|---|---|---|
| **No of units 4** | **Type** Single stage double suction | **Orientation** Horizontal |

| | **Units** | |
|---|---|---|
| **Fluid** | | Liquefied (partially hydrolysed starch) solution |
| | | |
| **Density** | **kg/m$^3$** | 985 |
| **Viscosity** | **Pa s** | 0.00031 |
| **Temperature** | **Deg C** | 90 |
| **Vapour pressure** | **Pa** | 70183 |
| **Solids (type)** | | NA |
| **Solids %** | | NA |
| **Flow rate** | **m$^3$/s** | 0.338 |
| **Total suction head** | **m** | 20.24 |
| **Total static head** | **m** | 3.10 |
| **Dynamic head** | **m** | 6.47 |
| **Total Dynamic head** | **m** | 9.58 |
| **Est. Efficiency** | | 0.75 |
| **Est. Power** | **kW** | 41.6 |
| **NPSH** | **m** | 1.82 |
| **Hazards** | | Irritation of skin, eyes, and respiratory tract in case of physical contact upon leakage. High operating temperature |
| **Materials** | | Carbon steel |
| **Seal** | | Hydrodynamic mechanical seal |
| **Seal liquid** | | Unbalanced mechanical seal |
| **Driver** | | Electric motor (AC supply) |
| **Drive speed** | **rpm** | 1200 |

| **Remarks** | | | | |
|---|---|---|---|---|
| **Rev no 1** | | | | |
| **Eng/date** | | | | |

## E-305 Centrifugal Pump Data Sheet

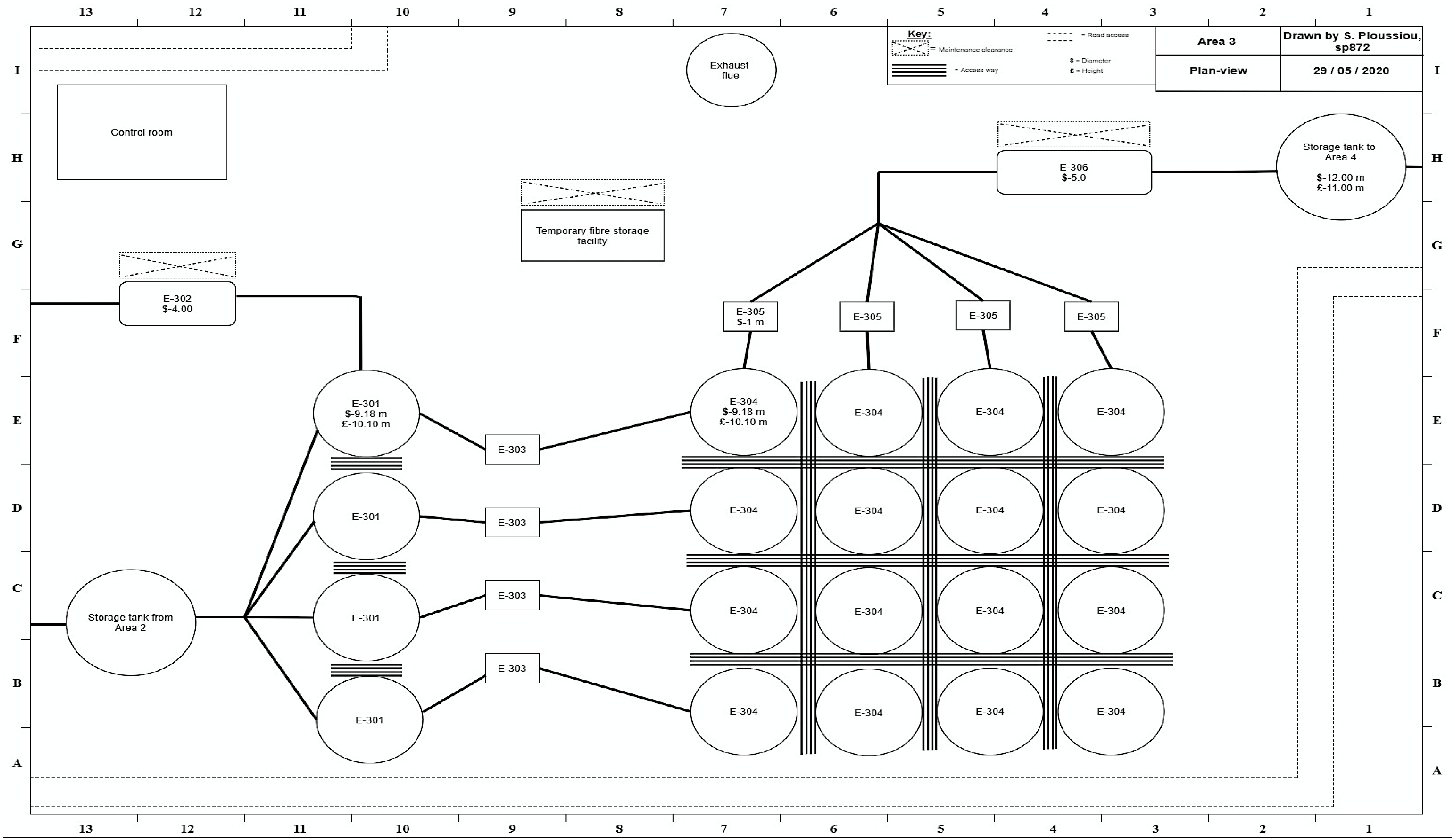
| | | |
|---|---|---|
| **Client** CEB | **Area** 3 | **Item no. 4** |

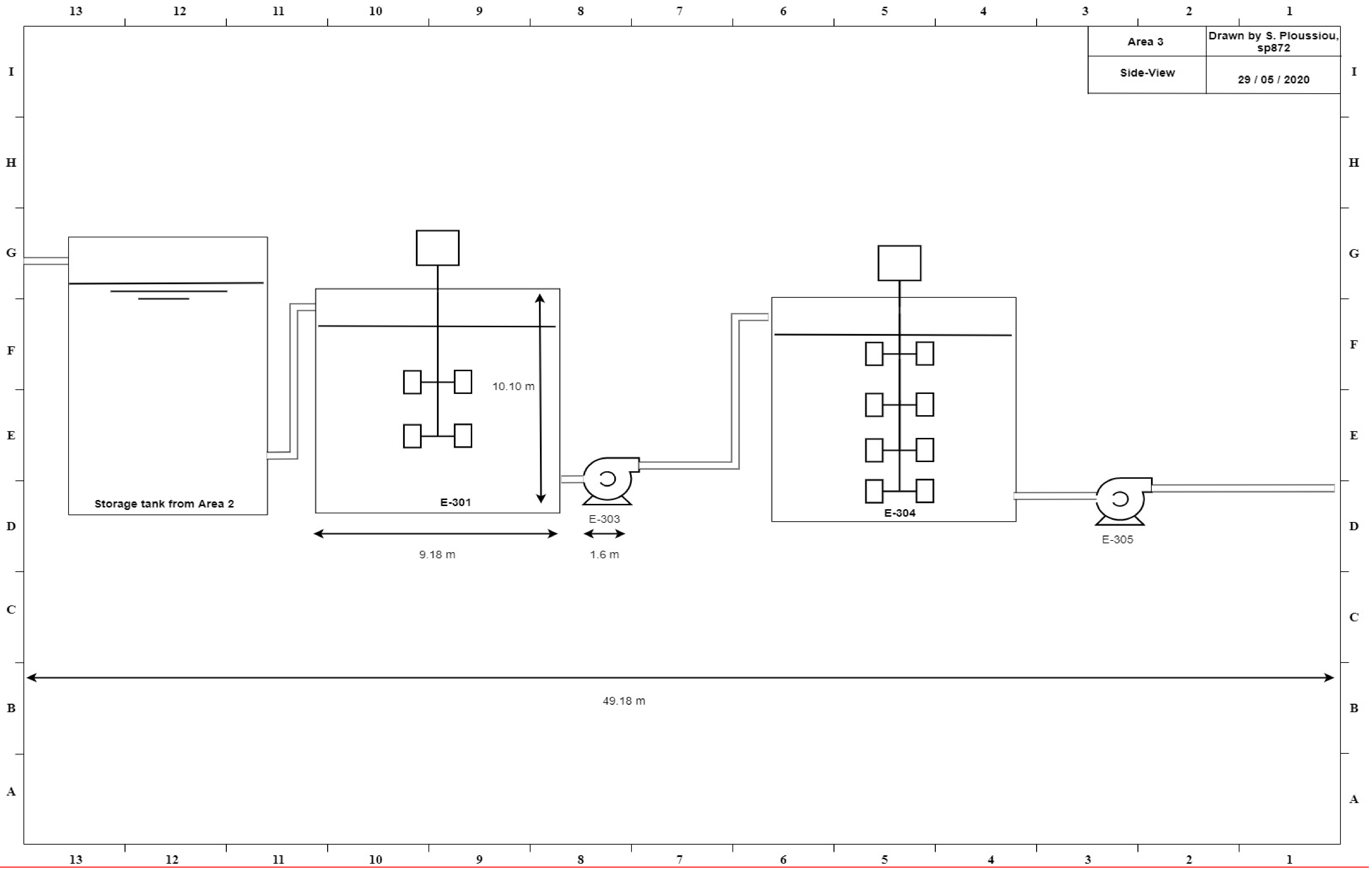
| | | |
|---|---|---|
| **Service** | | |
| **No of units 4** | **Type** Single stage double suction | **Orientation** Horizontal |

| | **Units** | |
|---|---|---|
| **Fluid** | | Dilute glucose solution |
| | | |
| **Density** | **$kg/m^3$** | 985 |
| **Viscosity** | **Pa s** | 0.00054 |
| **Temperature** | **Deg C** | 55.0 |
| **Vapour pressure** | **Pa** | 12000 |
| **Solids (type)** | | NA |
| **Solids %** | | NA |
| **Flow rate** | **$m^3/s$** | 0.338 |
| **Total suction head** | **m** | 19.53 |
| **Total static head** | **m** | 6.13 |
| **Dynamic head** | **m** | 12.94 |
| **Total dynamic head** | **m** | 19.08 |
| **Est. Efficiency** | | 0.75 |
| **Est. Power** | **kW** | 83.0 |
| **NPSH** | **m** | 18.96 |
| **Hazards** | | Mildly corrosive liquid upon leakage. Moderately high temperature |
| **Materials** | | Carbon steel |
| **Seal** | | Hydrodynamic mechanical seal |
| **Seal liquid** | | Unbalanced mechanical seal |
| **Driver** | | Electric motor (AC supply) |
| **Drive speed** | **rpm** | 1200 |

| | | | | |
|---|---|---|---|---|
| **Remarks** | | | | |
| **Rev no 1** | | | | |
| **Eng/date** | | | | |
| **Checked/date** | | | | |
| **Approved/date** | | | | |

5.6 Plan and side-view elevation sketches of Area 3




Area 3
Drawn by S. Ploussiou, sp872
Side-View
29 / 05 / 2020
Storage tank from Area 2
10.10 m
E-301
9.18 m
E-303
1.6 m
E-304
E-305
49.18 m

# 6 References

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