Australia's National
Science Agency

Evaluation of options for production of low cost CO_2-free hydrogen from Victorian brown coal

Final Report

Report number EP196700)
18 September 2019

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A Technical Report for Australian Carbon Innovation (formerly Brown Coal Innovation Australia Limited)

[1]: CSIRO; [2]: Monash University; [3]: Gamma Energy Technologies

CSIRO Energy

Citation

Lim S, Hayward J, Bhattacharya S, Bongers G (2019) “Evaluation of options for production of low cost CO_2-free hydrogen from Victorian brown coal”. CSIRO report for Australian Carbon Innovation.

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Acknowledgments

Funding for this project was provided by Cleantech Energy Australia Pty Ltd and Australian Carbon Innovation (formerly Brown Coal Innovation Australia Limited), a private member-based company with funding contracts through Australian National Low Emissions Coal Research and Development Ltd (ANLEC R&D) and the Victorian State Government.

Executive summary

Cleantech Energy Australia (CEA; formerly Coal Energy Australia) is exploring opportunities for exploiting the Victorian brown coal resource to produce low cost CO_2-free hydrogen. This project was a result of a joint funding arrangement established between BCIA, ANLEC R&D and CEA. The project entitled "Evaluation of options for production of low-cost CO_2-free hydrogen from Victorian brown coal" was initiated in accordance with the Research Project Agreement under the BCIA-funded project framework, dated 2 August 2017 and varied on 14 June 2018 and 11 October 2018 accordingly.

For this work, CSIRO was commissioned to conduct and coordinate a desktop costing study of potential CO_2-free hydrogen production technologies, in collaboration with Monash University and Gamma Energy Technology. The Part 1 of this report provides an overall summary of the study outcomes which were derived from three separate technical studies by the various project partners. The details of the respective studies led by various co-authors are also included in the report under Part II, III and IV respectively.

The study provided comparative assessments of various hydrogen production technologies based on pyrolysis and gasification of Victorian brown coal with CCS and water electrolysis using renewable (wind and solar) energy. The comparative assessment provided a good understanding of the relative competitiveness of the various proposed technology options from process economic and carbon emission considerations. The study was conducted using a common H_2 production basis of 770 tpd. The information will be used, in part, to guide any future project development and investment opportunity.

Based on the assumptions used in the current study, the gasification option was found to result in the lowest cost of hydrogen production, estimated as $2.73 – 4.64/kg. This is predominantly due to the lower CAPEX of this production route. The pyrolysis approach was generally shown to be higher cost than gasification - predominantly due to the fact that brown coal must be briquetted prior to pyrolysis, which is not required for gasification. The levelised costs of hydrogen production from renewable energy such as wind and solar in Victoria at 2030 were estimated to be at least 2 times higher than the gasification approach based on the chosen assumptions such as capacity factor, efficiency etc. For comparison purposes, the levelised cost based on Victorian's condition was also compared with that estimated from the National Hydrogen Roadmap study.

It should be acknowledged that estimated levelised costs of hydrogen production from this study are invariably dependent on the assumptions used in the current report which are outlined in the individual detailed studies presented in dedicated parts of the report. Readers are also encouraged to review the assumptions carefully with clear understanding of the context when making reference to this work. Ongoing peer-reviewed refinement of hydrogen production cost estimates is thus highly recommended, to reflect improved understanding of the changing cost structure facilitated by improvements to technology and availability of other relevant infrastructures, such as hydrogen storage and transport options.

Part I Overall Project Summary

Low cost CO_2-free hydrogen from Victorian brown coal - Evaluation of production options

1 Introduction

Around the world, increasing attention is being given to the potential of hydrogen to play an important role in global decarbonisation efforts. Hydrogen offers an alternative, zero emission, energy option as part of a diverse future global energy market. A variety of government-funded research and developments projects have been initiated to support the use of hydrogen as a source of energy.

Hydrogen offers a potential solution for the large-scale export of low or zero emissions energy. In order to do so, the hydrogen must be "CO_2-free". In this project, the term "CO_2-free hydrogen" includes zero emissions hydrogen produced from renewable energy via electrolysis, and low emissions hydrogen produced from fossil fuels, such as through pyrolysis or gasification of coal or biomass, or steam methane reforming of natural gas, in conjunction with carbon capture and storage (CCS).

The global interest in hydrogen is due to the growing pressure on countries to reduce their greenhouse gas emissions to mitigate the risks of climate change. The development of a 'hydrogen economy' is being led predominantly by Japan, in response to its Paris Agreement commitment to reducing greenhouse gas emissions by 26% before 2030.

A recent analysis by ACIL Allen Consulting forecast that, by 2030, the global market for CO_2-free hydrogen will be in the order of 3.8 million tonnes, worth about $9.5 billion. Of this, Australia could potentially supply 242,000 – 1,088,000 tonnes, generating $1.1 – 4.8 billion in export revenue.[1] Victoria is well positioned to play a key role in this market.

Victoria is in a unique position to become an important part of the value chain, given its abundant, world class brown coal resources in conjunction with the potential availability of large scale CO_2 storage site in the Gippsland Basin. The Japanese Institute of Applied Energy has estimated that CO_2-free hydrogen from Victoria may potentially contribute to about 30% of Japan's energy mix by the late 2030s. To meet this need, it is estimated that around 500,000 tonnes/year of hydrogen will need to be produced and exported from Victoria.[2]

The production of hydrogen by gasification of Victorian brown coal is currently being developed in the Hydrogen Energy Supply Chain project[3], led by Kawasaki Heavy Industries (KHI) and its industrial partners, with funding support from the Australian and Victorian governments.

Cleantech Energy Australia (CEA; formerly Coal Energy Australia) is also exploring opportunities for exploiting the Victorian brown coal resource to produce low cost CO_2-free hydrogen. This current project was a result of a joint funding arrangement established between BCIA, ANLEC R&D and CEA. CSIRO was commissioned to conduct a desktop costing study of potential CO_2-free hydrogen production technologies, in collaboration with Monash University and Gamma Energy Technology. The information from this study will, in part, guide any future project development and investment opportunity. This report provides an overall summary of the study outcomes.

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[1] Opportunities for Australia from Hydrogen Exports, ACIL Allen Consulting for ARENA, 2018.

[2] Kamiya, S., Nishimura, M., & Harada, E. (2015). Study on introduction of CO_2 free energy to Japan with liquid hydrogen. *Physics Procedia*, *67*, 11-19.

[3] https://hydrogenenergysupplychain.com/

2 Objective

The primary objective of the study was to conduct techno-economic assessments of a selected number of potentially low-cost CO_2-free hydrogen production technology options based on pyrolysis route from Victorian brown coal in order to guide future project development and investment opportunity.

The study included comparative technology assessments for brown coal pyrolysis approach with two other hydrogen production technologies, namely, brown coal gasification with CCS and electrolysis from renewable (wind and solar) energy based in Victoria.

The purpose of the comparative assessment is to understand the relative competitiveness of the various proposed technology options from process economic and carbon emission considerations. The competitiveness of these options would depend on many factors such as energy efficiency, carbon capture intensity, economics (OPEX, CAPEX) and other constraints.

3 Overall Approach

This project was conducted in two phases, where:

- Phase 1 involved development of process flow diagrams for the different processing options, mass and energy balance calculations using process modelling tool, identification and engagement with external consultants for an independent cost estimation study in Phase 2.
- Phase 2 involved cost estimation, compilation of a final report and presentation of the findings to the project stakeholders.

The project was managed and coordinated by CSIRO, with contributions from all project participants such as CEA, Monash University, Gamma Energy Technologies and BCIA. CEA provided business context and early stage project framing and definition. CSIRO and Monash University jointly formulated the process flowsheets for the various CO_2-free hydrogen production technology options based on brown coal resource i.e. pyrolysis and gasification as well as electrolysis. Specifics of the process options are summarised in Table 1.

Table 1. Different hydrogen production options used in the case studies.

Option	Process	Detailed Description
1	Pyrolysis gas only	Direct extraction of hydrogen from brown coal pyrolysis gas following a series of gas separation and purification processes
2	Pyrolysis and reforming	Production of hydrogen from brown coal pyrolysis gas using catalytic steam reforming approach and then followed by separation and purification techniques
3	Pyrolysis, tar cracking and reforming	Production of hydrogen from cracking of brown coal tar and pyrolysis gas using chemical looping-type reformer and then followed by gas separation and purification techniques
4	Gasification	Production of hydrogen from brown coal gasification using oxygen blown entrained flow gasifier and then followed by shift reactor and gas separation and purification techniques
5	Electrolysis with renewable energy (wind or solar)	Production of hydrogen using electrolysis process that is powered by renewable energy (wind or solar)

Option 1 was regarded as the simplest approach, similar to a typical hydrogen recovery system used in coke ovens in Japan. For Options 2 and 3, additional process units are incorporated to further convert the residual tar and volatile gases into hydrogen. Option 4 is based on typical gasification route where the produced syngas is further converted to hydrogen through a water-gas shift reaction. To achieve CO_2-free hydrogen production, all the process configurations based on brown coal route incorporated CO_2 capture and sequestration (CCS). Figure 1 shows the process block diagrams for the various hydrogen production options.

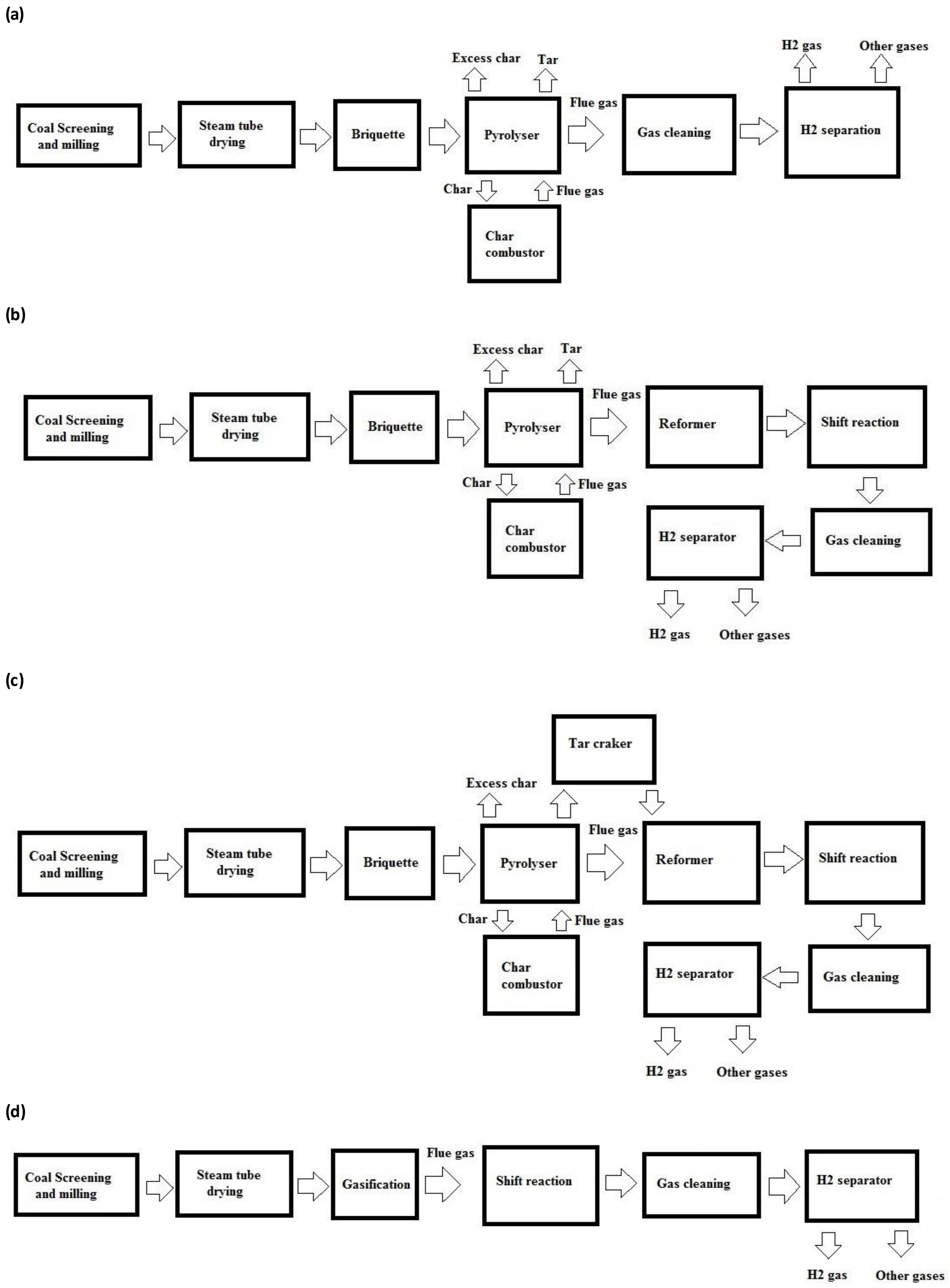

Figure 1. Process block diagrams for the various hydrogen production options based on brown coal. (a) Option 1 - Pyrolysis only; (b) Option 2- Pyrolysis and reforming; (c) Option 3 - Pyrolysis and tar cracking and reforming; (d) Option 4 - Gasification.

All the process options considered in the case studies included gas separation and purification techniques, with a focus on pressure swing adsorption (PSA) for hydrogen extraction and CO_2 capture based on calcium-based looping. $CaCO_3$ looping was selected as a promising CO_2 capture technique that employs a widely available and inexpensive sorbent and does not require a fresh water stream like conventional amine scrubbing.

A series of process simulations were conducted by Monash University using the a process simulation package, in order to determine the mass and energy balance for each process. The input data were sourced from CEA as well as from available literature. To enable a common and meaningful basis for comparison, the techno-economic case studies were conducted for nominal hydrogen production capacity of 770 tonnes/day (225,500 tonnes/year @~80% utilisation), based on an early published study developed for the Hydrogen Energy Supply Chain project.[4]

The simulation study enabled the quantification of key process parameters such as the amount of brown coal required, the anticipated CO_2 emissions, make-up water and heat requirements and amount of residual char for the various process options. The material and energy balances from the process simulations were used in a subsequent costing study.

A comprehensive cost analysis study was then conducted in order to estimate the levelised costs of hydrogen production from the various hydrogen production options based on brown coal. Gamma Energy Technology adopted the costing methodology outlined by the Association for Advancement of Cost Engineering International (AACEI). The cost analysis study considered estimation of process OPEX, CAPEX and other associated costs.

To provide a comparison with competing CO_2-free H_2 production technologies based on renewable energy, a separate cost analysis was also conducted by CSIRO (shown as Option 5 in Table 1). The Option 5 considers hydrogen production process based on polymer electrolyte membrane (PEM) electrolysis, powered by renewable energy (solar PV or wind). To allow direct comparison with the coal-based options, the PEM process was scaled to allow a hydrogen production capacity of 770 tonnes/day in Victoria. Since this scale of production could not be accommodated using excess grid-based electricity, the cost of dedicated renewable energy infrastructure was included in the cost estimates.

To develop appropriate costs for renewable hydrogen production, CSIRO adapted the methodology used in its National Hydrogen Roadmap study[5] to reflect the Victorian context and conditions. Only direct hydrogen production costs were included in this study. Additional cost factors, such as transportation of product to an export hub and other infrastructure costs, were not included.

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[4] Kamiya, S., Nishimura, M., & Harada, E. (2015). Study on introduction of CO_2 free energy to Japan with liquid hydrogen. *Physics Procedia*, *67*, 11-19.

[5] https://www.csiro.au/en/Do-business/Futures/Reports/Hydrogen-Roadmap

4 Overall project findings

This study has produced a comprehensive comparison of hydrogen production costs – based on brown coal and renewable resources. The input parameters and approaches used were specifically tailored for the Victorian context, e.g. utilisation technology options for Victorian brown coal and availability of renewable energy resources within the state (i.e. not confined to the Latrobe Valley). The study considered the scale of processing plants (coal feed, CO_2 emission intensity), water consumption and land use required to meet a target hydrogen production capacity of 770 tonnes/day. For the brown coal options, it was assumed that the technologies for carbon capture and sequestration (CCS) would be readily available at commercial scales in 2030 for mitigation of CO_2 emissions.

As part of the project, three separate supplementary technical studies were conducted by the project participants. These are described in detail in the various parts of this report, namely:

Part II: "Process simulation of hydrogen production options from Victorian brown coal", by M. Kibria and S. Bhattacharya, Monash University
Part III: "Cost analysis of hydrogen production from Victorian brown coal", by G. Bongers, S. Byrom and Z. Pregelj, Gamma Energy Technology and Altaprom International.
Part IV: "The economics of producing hydrogen from electrolysis in Victoria", by J. Haywood, CSIRO Energy.

The key findings and highlights are summarised below, and the full reports are provided as appendices.

4.1 Hydrogen from brown coal resources

Table 1 below summarises the key process parameters for production of 770 t/d hydrogen using the three pyrolysis options and gasification. The data in Table 2 is presented on an annual basis, calculated by multiplying the estimated hourly rates by 7074 hour/year (based on 85% availability and 95% utilisation).

Table 2. Key process parameters for the production of 770 t/d (~225,500 t/y) of hydrogen.

	Process Options			
	1	2	3	4
Raw coal (Mt/y)	26.0	12.6	10.5	6.3
CO_2 generated (Mt/y)	9.8	6.5	6.4	3.7
Make-up process water needed (GL/y)	1.2	2.6	2.4	2.4

- The total amount of brown coal required to achieve the 770 t/d of hydrogen was the largest for Option 1, and the quantity decreased progressively from Options 2 to 4. The gasification route required the least amount of brown coal input, as the process is inherently maximised for syngas (thus hydrogen) production, whereas the pyrolysis process was designed for production of both char and pyrolysis gas.
- The amount of brown coal required for the gasification route, 6.3 million t/y, is significantly less than currently mined at either Yallourn (18 million t/y)[6] or Loy Yang (30 million t/y)[7].
- The gasification route produced the least amount of carbon dioxide, estimated at 3.7 Mt/y. For comparison, CarbonNet has determined that the proposed Pelican storage site in the Gippsland Basin has at least 125 million tonnes of CO2 storage capacity[8], sufficient for 33 years of hydrogen production.
- The amount of make-up process water required for gasification was much the same as that needed for the best pyrolysis option (Option 3). In each case, the amount of make-up water required is less than the amount of water evaporated during drying of the coal. However, the costs associated with recovery of water from dryer condensate were not considered in this report.

Table 3 below shows the forecast capital costs (CAPEX) and levelised cost of product (LCoP) in 2030.

Table 3. Estimated CAPEX and LCoP in 2030

Option	Production type	CAPEX ($M)	LCoP ($/kg)*
1	Pyrolysis gas only	16,578 – 30,946	9.49 – 16.86
2	Pyrolysis and reforming	8,460 – 15,794	4.47 – 8.61
3	Pyrolysis, tar cracking and reforming	7,519 – 14,036	3.78 – 7.19
4	Gasification	4,318 – 8,052	2.73 – 4.64

** Note: The LCoP estimates do not include the costs associated with mine remediation at the end of production.*

- Among the different pyrolysis based approaches, Options 2 and 3 offer slightly better process economics than Option 1, because the energy contained in the fuel gas is used to extract hydrogen from water. It should be noted that the technologies involved in Options 2 and 3 are not commercially available, and would require additional research and development to increase their technology readiness level.
- The gasification option was found to require the lowest capital expenditure. The higher cost of pyrolysis was predominantly due to the fact that brown coal must be briquetted prior to pyrolysis, which is not required for gasification. This made the proposed pyrolysis process options less competitive when compared to the gasification process.
- The gasification option was also found to result in the lowest cost of hydrogen production, estimated as $2.73 – 4.64/kg. This is predominantly due to the lower CAPEX of this production route. It should be noted that the cost estimation is focused on the production of hydrogen at source but did not consider the longer term of cost of mine site rehabilitation after the end of mine life.

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[6] https://www.energyaustralia.com.au/about-us/energy-generation/yallourn-power-station.

[7] https://www.bioregionalassessments.gov.au/assessments/12-resource-assessment-gippsland-basin-bioregion/12215-loy-yang-mine-and-power-station

[8] https://earthresources.createsend1.com/t/ViewEmail/r/CA00E014249083542540EF23F30FEDED

4.2 Hydrogen from renewable resources

For comparison purposes, a separate costing study was also carried out to estimate the levelised cost of hydrogen production ($LCOH_2$) from renewable energy. The cost estimates exclude other cost factors such as transportation of product to the market or end-user and other infrastructure costs.

- In order to meet the requirement for 770 t/d of hydrogen, several large-scale renewable energy farms and electrolysis systems would need to be constructed. It was found that up to approximately 7 GW of wind or 9.4 GW of photovoltaics (PV) would be conservatively required along with a similar capacity of PEM electrolysers. For costing purposes, it was assumed that the individual systems would be of the order of ~500 MW and that several of each type would be built across the whole of Victoria for hydrogen production at close to the energy source.
- The study estimated the $LCOH_2$ for the years 2017, 2020 and 2030 based on certain assumptions of technology cost and performance improvement factors over the specified period. The estimated $LCOH_2$ for wind- and PV-powered electrolysis projected at 2030 are included in Table 3 above.
- For both energy options, the cost was found to vary from \$5.6/kg H_2 to \$8.9/kg H_2 over the forecast period up to 2030. The values were similar because of the relatively high cost of electrolysis. Wind power was found to be somewhat lower in cost than the PV, based on the current assumptions. This was due to the better availability of wind resources than solar in Victoria. However, the gap of production cost between wind and PV becomes narrower in later years, reflecting an anticipated fall in the capital cost of PV due to technology improvement.
- The study also evaluated the extent of resource use that would be required for production of export quantities of hydrogen by renewable-powered electrolysis.
- Production of a nominal capacity of 770 tonnes/day (225,500 tonnes/year – assumed at 80% utilisation factor) of hydrogen by electrolysis of water would consume around 2 billion litres/year of purified water. The availability of suitable sites for renewable hydrogen production in term of energy and water resources will need careful consideration.
- There is no uniform definition of the total area of a wind power plant. A survey of wind farms in the United States[9] suggested that there are two primary indices of land use – the infrastructure/direct impact area (or land temporarily or permanently disturbed by wind power plant development) and the total area (or overall area of the power plant as a whole). For 93 projects representing about 14 GW of proposed or installed capacity, the average permanent direct impact value reported was 0.3 ± 0.3 hectares/MW of capacity. For 161 projects representing about 25 GW of proposed or installed capacity, the average value for the total project area was about 34 ± 22 hectares/MW of capacity. Our analysis of wind farm area data in Victoria indicated that the average total wind farm area is 31.62 ± 17.44 ha/MW, which is consistent with the US wind farm data.
- On the basis of the more credible US data, the total estimated wind farm area required in 2030 to generate 7 GW is 238,000 ± 154,000 ha, with a permanent direct impact area of 2,100 ± 2,100

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[9] Denholm, P., Hand, M., Jackson, M., & Ong, S. (2009). Land-Use Requirements of Modern Wind Power Plants in the United States. NREL.

ha. For illustrative purposes, the largest capacity wind farm in Victoria, the 420 MW McArthur wind farm in Victoria, occupies a total area of 5,500 ha. Production of 770 t/d hydrogen using wind will thus require the equivalent area of 43 ± 28 (or 15 - 71) times McArthur wind farms. Note that the majority of this land could be used for other purposes such as grazing.

- For PV, production of 9.4 GW electricity will require 15,000 ha of bare panels, or 23,000 ha installed. For illustrative purposes, the largest PV farm in Australia is the 102 MW Nyngan Solar Plant in NSW, which occupies an area of 250 ha. Production of 9.4 GW electricity by PV will require an area equal to 92 equivalents of the Nyngan Solar Plant.
- From this it may be concluded that production of export quantities of hydrogen by renewable energy will likely be limited by the large areas of land required. The widely dispersed nature of renewable hydrogen production would increase the transport and handling costs, but these were not included in the figures shown in Table 3.

5 Concluding remarks

This project has enabled an objective evaluation of technical and economic viability of the various options for producing CO_2-free hydrogen, including three brown coal pyrolysis-based processes, brown coal gasification, as well as electrolysis using renewable electricity from either wind or solar energy.

The cost analysis found that the projected levelised cost of hydrogen production in 2030 is lowest for the gasification option, at \$3.50/kg. This is somewhat higher than the cost estimated in the National Hydrogen Roadmap (Bruce et al 2018), which suggested a cost for brown coal gasification plus CCS of \$2.14–\$2.62/kg once the commercial scale production and CCS plant come online in the 2030s.

ACIL Allen Consulting suggested that liquefaction, loading and shipping to Japan would add an extra \$2.11/kg to the cost. This suggests that the landed price in Japan of hydrogen produced by gasification would be \$5.61/kg. This is equivalent to ¥37/Nm^3 (assuming an exchange rate of ¥80/\$), which is close to the import price of ¥35/Nm^3 assumed by METI in its projections.[10] Gasification of brown coal plus CCS thus appears to be the only technology option capable of producing CO_2-free hydrogen at the cost needed to access the Japanese market.

The direct extraction of hydrogen from pyrolysis gas is not competitive due to the low concentration of hydrogen in pyrolysis gas and the corresponding large scale of the plant including briquetting required to achieve a capacity of 770 t/d of hydrogen. Other augmented pyrolysis process options (Options 2 and 3) with gas reformer or tar cracker may offer slightly improved economics than the simple pyrolysis option (Option 1). However, further research and development would be required to improve the technology readiness of the augmented reformer or tar cracker.

Based on the current cost analysis, the levelised costs of hydrogen production from renewable energy such as wind and solar at 2030 were estimated to be at least 2 times higher than the gasification approach. It was also found that the feasibility of producing hydrogen by electrolysis of water using wind- or solar-based renewable electricity is likely to be limited by the large areas of land required and the need for massive quantities of purified water in (likely) remote areas.

Consideration of the availability of suitable land and water, social and environmental impacts, and other influencing factors such as proximity to the market, transportation requirements and long-term mine rehabilitation, were beyond the scope of this study.

Ongoing peer-reviewed refinement of hydrogen production cost estimates is highly recommended, to reflect improved understanding of the changing cost structure facilitated by improvements to technology and availability of other relevant infrastructures, such as hydrogen storage and transport options. This will be important to assist both public policy development and in advancing commercial interests. Readers are also encouraged to review the assumptions carefully with clear understanding of the context when making reference to this work.

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[10] Kamiya, S., Nishimura, M., & Harada, E. (2015). Study on introduction of CO_2 free energy to Japan with liquid hydrogen. *Physics Procedia*, *67*, 11-19.

Part II Supplementary Report 1

Process simulation of hydrogen production options from Victorian brown coal

By

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6 Background

Hydrogen is considered an energy carrier; like electricity, it must be produced from a primary energy source such as biomass or hydrocarbons such as natural gas or coal. Four different thermal processing options for hydrogen production using Victorian brown coal are investigated in this report. All the studied processes are coupled with carbon capture facilities. The four options are summarised below:

Option 1	Brown coal pyrolysis plant for H_2 production
Option 2	Brown coal pyrolysis plant followed by shift reactor for H_2 production
Option 3	Brown coal pyrolysis plant followed by shift and tar cracking reactor for H_2 production
Option 4	Brown coal gasification plant using oxygen blown entrained flow gasifier followed by shift reactor for H_2 production.

7 Modelling approach

Commercially available process simulator is a valuable tool for design, operation, and evaluation of chemical process that consist of unit operations connected by process stream. By ensuring mass and energy balance for each unit operation as well as the overall process, the simulator can calculate unknown stream flowrates, temperature, pressures and compositions. The process simulation software includes built-in library models for the most common unit operations and more databanks of components with parameters and method to calculate thermodynamic properties.

7.1 Equation of state

The Base Property method (Equation of state) used for the simulation was the IDEAL with a COMMON filter.

7.2 Conventional components

These are internal components of the process modelling software for which the physical property correlations are already defined within the software databank (Table 4):

Table 4 Conventional components

Component ID	Type	Component name	Alias	State
H_2	Conventional	HYDROGEN	H_2	Gas
CO	Conventional	CARBON-MONOXIDE	CO	Gas
CO_2	Conventional	CARBON-DIOXIDE	CO_2	Gas
SO_2	Conventional	SULFUR-DIOXIDE	SO_2	Gas
SO_3	Conventional	SULFUR-TRIOXIDE	SO_3	Gas
NO_2	Conventional	NITROGEN-DIOXIDE	NO_2	Gas
NO	Conventional	NITRIC-OXIDE	NO	Gas
S	Solid	SULFUR	S	Solid
H_2O	Conventional	WATER	H_2O	Liquid
COAL	Nonconventional			

C	Solid	CARBON-GRAPHITE	C	Solid
C_2H_6	Conventional	ETHANE	C_2H_6	Gas
CH_4	Conventional	METHANE	CH_4	Gas
H_2S	Conventional	HYDROGEN-SULFIDE	H_2S	Gas
C_3H_8	Conventional	PROPANE	C_3H_8	Gas
C_3H_6-2	Conventional	PROPYLENE	C_3H_6-2	Gas
O_2	Conventional	OXYGEN	O_2	Gas
N_2	Conventional	NITROGEN	N_2	Gas
NH_3	Conventional	AMMONIA	NH_3	Gas
$C_{10}H_8$	Conventional	NAPHTHALENE	$C_{10}H_8$	Liquid
C_6H_6O	Conventional	PHENOL	C_6H_6O	Liquid
TERT--01	Conventional	TERT-BUTYL-ACETATE	$C_6H_{12}O_2$	Liquid
C_2H_4	Conventional	ETHYLENE	C_2H_4	Gas
CHAR	Nonconventional			
P-CRE-01	Conventional	P-CRESOL	C_7H_8O-5	Liquid
ASH	Nonconventional			
CaO	Solid	CALCIUM-OXIDE	CaO	Solid
$CaCO_3$	Solid	CALCIUM-CARBONATE-CALCITE	$CaCO_3$	Solid
$CaSO_3$	Solid	CALCIUM-SULFITE	$CaSO_3$	Solid
$CaSO_4$	Solid	CALCIUM-SULFATE	$CaSO_4$	Solid
CaS	Solid	CALCIUM-SULFIDE	CaS	Solid
Fe	Solid	IRON	Fe	Solid
Fe_2O_3	Solid	HEMATITE	Fe_2O_3	Solid
FeO	Solid	FERROUS-OXIDE	FeO	Solid
Fe_3O_4	Solid	MAGNETITE	Fe_3O_4	Solid

7.3 Non-conventional components

COAL, CHAR and ASH are defined as a non-conventional component. Default IGT relation is used for density and Specific heat calculation.

7.4 Tar

Tar (Pyrolysis oil) product is represented as a combination of four conventional hydrocarbon components. These components were selected from experimental data and literature review.

7.5 Particle size distribution

Particle size distribution (PSD) was modelled in the feed preparation area. A PSD was defined using 25 sizes (mesh) with a top size of 75 mm. The Rosin-Rammler-Sperling-Bennett (RRSB) method with dispersion parameter of 1 and $^{*}D_{63}$ of 0.018 m was used to create the mass fraction of the coal (See Figure 2), where $^{*}D_{63}$ is the particle size that corresponds to 63% of the total volume distribution. The parameters were set to affect the PSD change accordance with over/undersize mass balance data provided by Coal Energy Australia (now Cleantech Energy Australia).

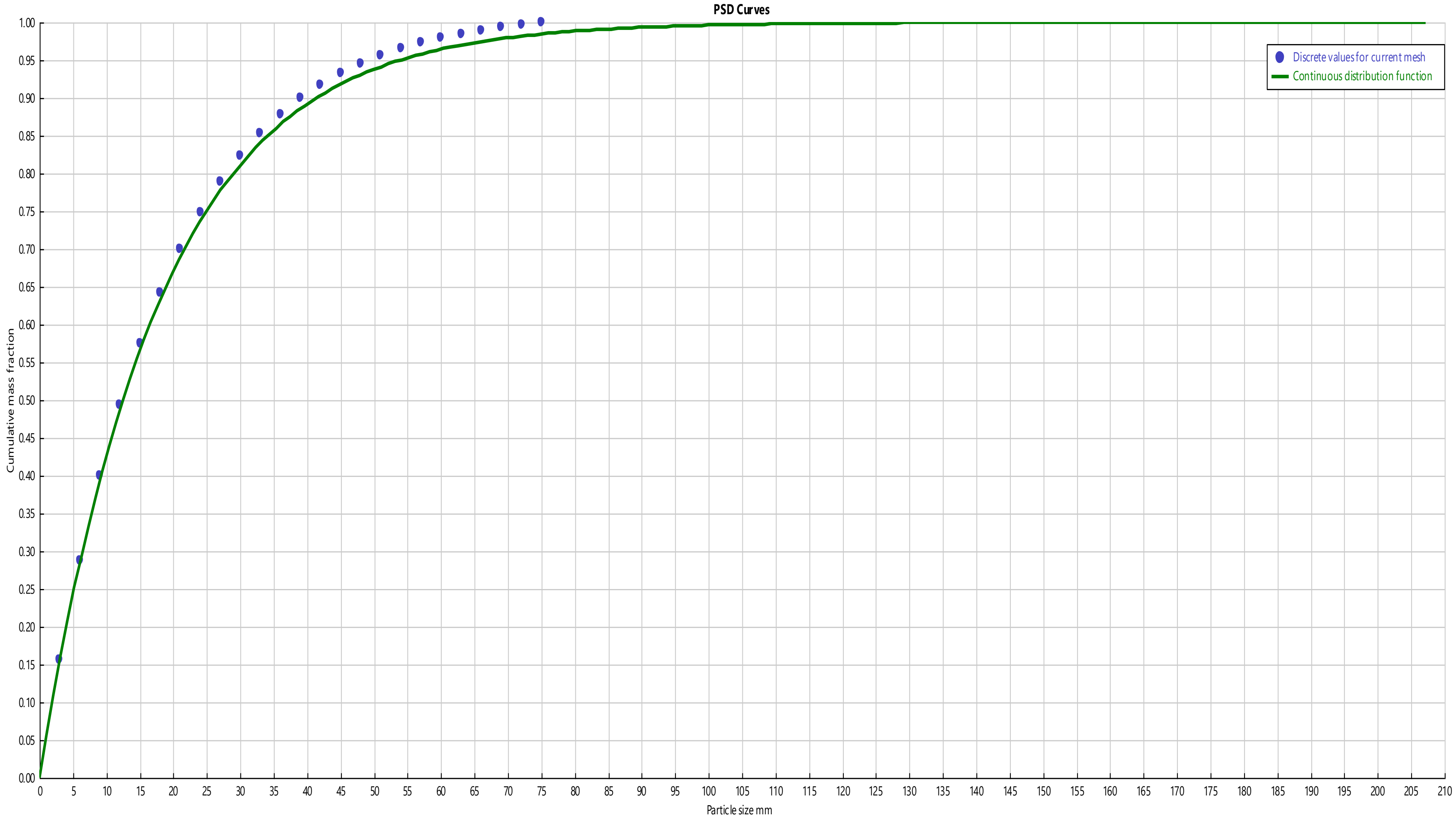

Figure 2: Coal PSD curve

8 Process description

8.1 Option-1: Brown coal pyrolysis plant for H_2 production

The simplified block diagram of option-1 is presented in Figure 3.

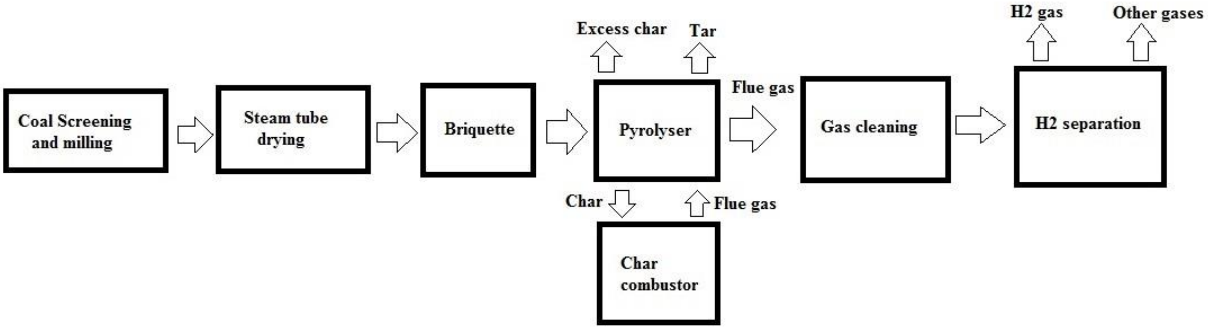

Figure 3: Block diagram of Option-1

The raw coal is processed by subsequent two-step hammering and screen to the desired particle size for efficient drying. The wet coal loses almost 55% of its initial weight during the drying process. The dry coal then goes to a briquette unit in which the dry coal is compressed to cylindrical pellets. These pellets are then fed to a Pyrolyser unit. A certain amount of char is combusted to produce the heat energy to run the pyrolyser and to produce steam for the steam tube dryer.

The flue gas is then quenched at low temperature to separate the Tar and subsequently passed to a Calcium carbonate looping system for CO_2 separation. This is chosen over the conventional amine process as it is assumed the higher efficiency carbonate looping system will be available by 2030. The CO_2 free gases then moved through a pressure swing adsorption process to separate the Hydrogen.

Key results obtained:

Target production of H_2	32.1 tons/hour
Wet coal requirement	3680 tons/hour
Dry coal	1136 tons/hour
Excess char to refinery	69 tons/hour
Tar production	124 tons/hour

Total CO_2 generation	1391 tons/hour
CO_2 capture efficiency	86%
Steam requirement	173 tons/hour

8.2 Option-2: Brown coal pyrolysis plant followed by shift reactor for H_2 production

Option- 2 is similar to option-1, except this option has a water gas shift unit coupled with Methane steam reformer unit.

The flue gas leaving the pyrolyser unit is rich in Carbon-monoxide (CO) and Methane (CH_4). To increase the H_2 production level, the available Methane is transformed to H_2 and CO in the Methane steam reformer unit shown in Eq. (1).

$$CH_4+H_2O \leftrightarrow CO+3H_2 \quad (1)$$

The CO-rich flue gas is further passed through a water gas shift reactor to boost up the H_2 production level based on Eq. (2)

$$CO+H_2O \leftrightarrow CO_2+H_2 \quad (2)$$

The block diagram of the process is presented in Figure 4.

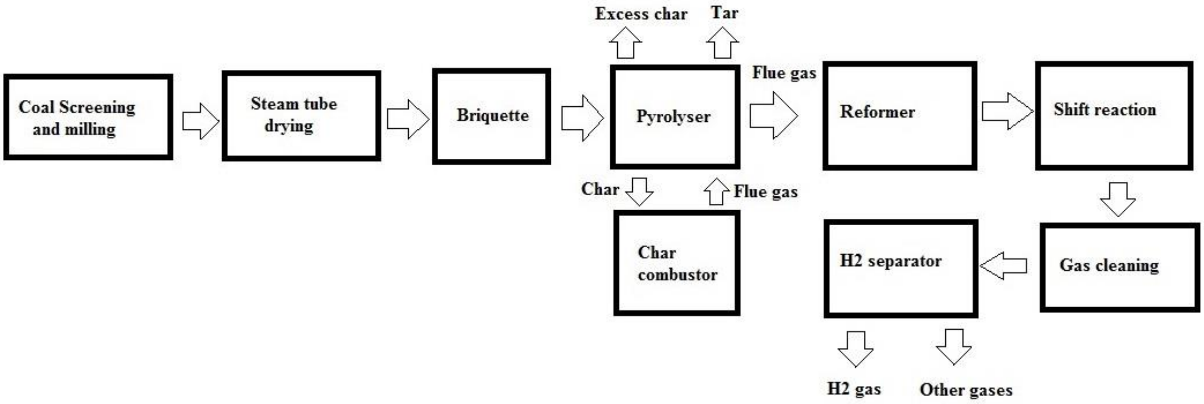

Figure 4: Block diagram of Option-2

Key results obtained:

Target production of H_2	32.1 tons/hour
Wet coal requirement	1780 tons/hour

Dry coal	550 tons/hour
Excess char to refinery	83 tons/hour
Tar production	62 tons/hour
Total CO_2 generation	924 tons/hour
CO_2 capture efficiency	93%
Steam requirement	375 tons/hour

8.3 Option-3: Brown coal pyrolysis plant followed by shift and tar cracking reactor for H_2 production

Option- 3 is similar to option-2, except this option has a Fe_2O_3 looping process.

The quenched and cooled flue gas after pyrolysis has a significant amount of Tar (heavy-oil), which goes in a Tar cracking unit. In this unit, the long chain hydrocarbon is broken down to light hydrocarbon and passed through water gas shift reactor followed by the Methane steam reforming reactor. The block diagram of the process is presented in Figure 5.

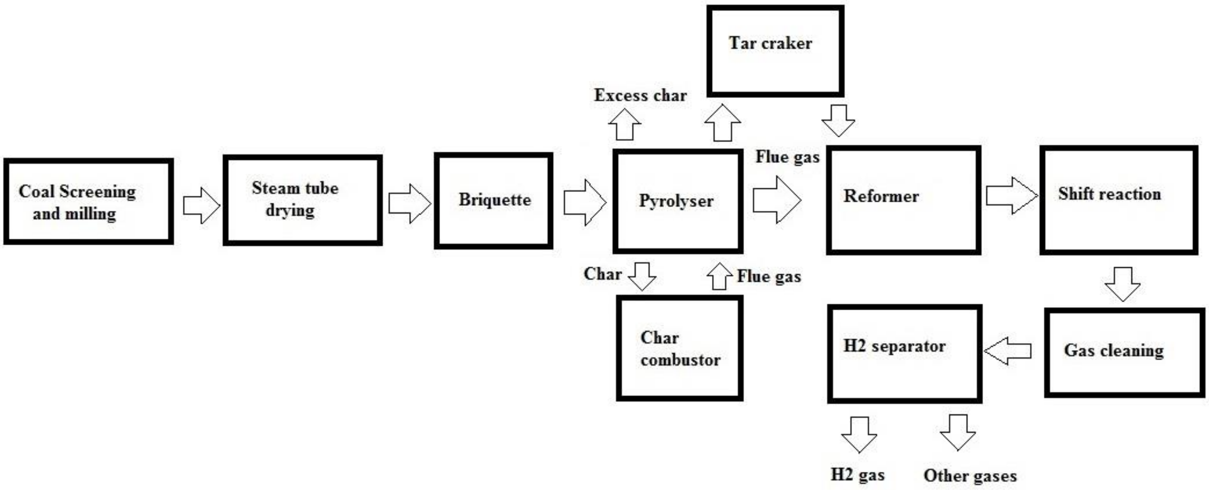

Figure 5: Block diagram of Option-3

Key results obtained:

Target production of H_2	32.1 tons/hour
Wet coal requirement	1480 tons/hour
Dry coal	457 tons/hour
Excess char to refinery	67 tons/hour
Tar production	Nil

Total CO_2 generation	902 tons/hour
CO_2 capture efficiency	94%
Steam requirement	340 tons/hour

8.4 Option-4: Brown coal gasification plant using oxygen blown entrained flow gasifier followed by shift reactor for H_2 production

The raw coal is processed by subsequent two-step hammering, screen and drying process to the desired particle size before feeding to the entrained flow gasifier. The entrained flow gasifier used in the simulation is a two-stage oxy-fired slagging gasifier which mimics the Mitsubishi Heavy Industry (MHI) gasifier. The syngas after gasifier unit is CO-rich. The syngas is cooled and subsequently passed through a shift reactor to increase H_2 level. The syngas goes to a Calcium carbonate looping system for CO_2 separation. The CO_2 free gases then passed through a pressure swing adsorption process to separate the Hydrogen. The block diagram is presented in Figure 6.

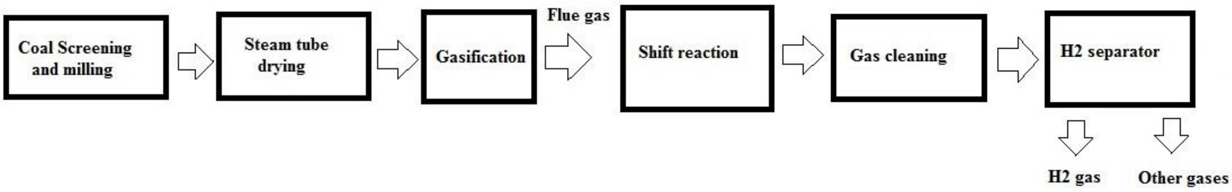

Figure 6: Block diagram of Option-4

Key results obtained:

Target production of H_2	32.1 tons/hour
Wet coal requirement	893 tons/hour
Dry coal	332 tons/hour
Excess char to refinery	Nil
Tar production	Nil
Total CO_2 generation	518 tons/hour
CO_2 capture efficiency	88%
Steam requirement	344 tons/hour

8.5 Separate case study for 18 million tons/year of wet coal

For comparison purposes, a separate case study is carried out to estimate the net H_2 production through a pyrolysis plant based on the present mining capacity of 18 million tonnes of raw coal per annum from Yallourn's open cut mine, Victoria Australia[1]. The process diagram is similar to option-1. However, in this case 2055 tons/hour of wet coal equivalent to 18 million tons/year is processed to estimate the generation of hydrogen which is around 17 tons/hour as opposed to 32.1 tons/hour. This comparison is intended to provide readers a sense of the scale of the coal mining and production using Yallourn mine as reference point.

Key results obtained:

Target feed wet coal	2055 tons/hour
Dry coal	634 tons/hour
H2 production	17 tons/hour
Excess char to refinery	80 tons/hour
Tar production	71 tons/hour
Total CO_2 generation	657 tons/hour
CO_2 capture efficiency	91%
Steam requirement	96 tons/hour

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[1] https://www.energyaustralia.com.au/about-us/energy-generation/yallourn-power-station

9 Overall findings

This study has explored the pathways of Hydrogen (H_2) production from Victorian brown coal (Yallourn coal) through a thermal process with CO_2 separation facilities. The target production of H_2 was set at 770 tons/day. The following four different options were investigated:

Option 1 Brown coal pyrolysis plant for H_2 production

Option 2 Brown coal pyrolysis plant followed by shift reactor for H_2 production

Option 3 Brown coal pyrolysis plant followed by shift and tar cracking reactor for H_2 production

Option 4 Brown coal gasification plant using oxygen blown entrained flow gasifier followed by shift reactor for H_2 production.

The steady-state thermodynamic simulation was performed using process simulation software package. All the options indicated were modelled with CO_2 separation facilities through $CaCO_3$ looping including upstream feed preparation, steam tube drying of coal, briquette preparation for pyrolysis, pyrolysis reactor, oxygen blown entrained flow gasifier, water gas shift and methane steam reformer reactor and associated downstream gas processing units including heat and materials balance.

The result of H_2 production (32.1 tons/hour equivalent to 770 tons/day) from the options listed above is presented in Table 5:

Table 5 Comparative results for coal consumption (Tons/hr)

	Wet coal	Dry coal
Option 1	3680	1136
Option 2	1780	550
Option 3	1480	457
Option 4	893	332

Table 6 presents the comparative char and Tar productions among the options

Table 6 Comparative results of char generation, char combusted and tar production (Tons/hr)

	Char generation	Combusted	Excess char to refinery	Tar
Option 1	575	506	69	124
Option 2	278	195	83	62
Option 3	231	164	67	NA
Option 4	225	34 + 191**	NA	NA

**Gasified in the reduction zone of Entrained flow gasifier

Table 7 presents the net CO_2 production, separation and make up $CaCO_3$ requirements for the separation

Table 7 Comparative results CO_2 generation and separation (Tons/hr)

	Total CO_2 generation	Separated with $CaCO_3$ looping	To Flare	Separation efficiency (%)	Makeup $CaCO_3$ requirements
Option 1	1391	1200	191	86	315
Option 2	924	865	59	93	150
Option 3	902	850	52	94	98**
Option 4	518	457	61	88	80

** 80 tons/hour Fe_2O_3 required for tar cracking unit

Table 8 presents makeup water requirements for the options:

Table 8 Makeup water requirements (tons/hour)

	Dryer	Shift reactor	Steam reformer	Auxiliary	Total
Option 1	168	X	X	5	173
Option 2	81	191	100	3	375
Option 3	67	190	80	3	340
Option 4	37	307	x	3	344

As presented in Table 5, the lowest coal requirement for the same level of H_2 production (32.1 tons/hour) can be achieved from Option-4. Table 9 presents the ratio of coal requirements compared to Option -4.

Table 9 Ratio of wet coal requirements compared to OPTION-4

Option 1	4.12
Option 2	1.99
Option 3	1.65

As shown in Table 6, Option -2 has the most potentiality to produce excess Char along with Tar that might be products during the process. For more information on Tar (Heavy oils), please refer to the Appendix section. Table 10 presents the comparative table of char combustion for heat energy production to sustain the whole process.

Table 10 Comparative (%) of char combustion

Option 1	88
Option 2	70
Option 3	71
Option 4	15

As seen in Table 7, for 88% CO_2 capture efficiency, Option-4 requires the least makeup $CaCO_3$. However, during Option-3, the CO_2 capture efficiency can be maximized up to 94%. Option-4 produces the least CO_2 during the process which also reflects on Table 10 in terms of char combustion which is the primary cause of CO_2 generation. From Table 8**,** it is evident that total water requirements for Option-2, 3, 4 are almost similar. However, Option-1 requires the maximum makeup water for the dryer unit as this option requires the maximum wet coal for the same purpose.

Apart from that, a case study is also carried out to locate the net H_2 production through a pyrolysis plant with the present mining capacity of 18 million tons of raw coal from Yallourn's open cut mine, Victoria Australia. The results are presented in Table 11.

Table 11 H_2 production from 18 million tons/year of wet coal

Product of H_2	17 tons/hour
Wet coal	2055 tons/hour (Equivalent to 18 million tons/year)
Dry coal	634 tons/hour
Total char generation	321 tons/hour
Excess char to refinery	80 tons/hour
Tar production	71 tons/hour
Total CO_2 generation	657 tons/hour
CO_2 capture efficiency	91%
Steam requirement	96 tons/hour

For clarity of presentation, details of the simulation results produced for each of the Options 1, 2, 3 and 4 are provided in the Appendix section, at the end of this compilation report.

Part III Supplementary Report 2

Cost analysis of hydrogen production from Victorian brown coal

G. Bongers & S. Byrom – Gamma Energy Technology

Z. Pregelj - Altaprom International

10 Introduction

Interest in hydrogen as an energy carrier and an energy storage medium has increased in recent times.[2, 3] One of hydrogen's advantages is based on the possibility that almost any source of energy can be converted into hydrogen and it has a wide range of uses. This costing study focuses on four options developed by Monash University and CSIRO, all of which are based on Victorian brown coal.

Hydrogen offers an alternative energy option as part of a diverse future global energy market, with Australia, and specifically Victoria, in a unique position to become an important part of the hydrogen value chain. Australia along with other major advanced economies, and particularly Japan, are making investments into hydrogen production. Given Australia's large brown coal resources, Victoria is very well-placed to become a global, cost competitive producer of hydrogen. The conversion of brown coal into hydrogen offers one of the best ways to provide a steady and reliable energy source that is not dependent on the weather. As such investigations into a zero emission hydrogen solution using advanced carbon capture and storages technologies should be a priority.

This study was commissioned by CSIRO as part of its project, 'Options for low cost production of CO_2-free hydrogen', co-funded by CEA P/L and BCIA, a private member-based company with funding contracts through ANLEC R&D and the Victorian State Government. The precursor for this work was a process simulation study, by Monash University and CSIRO, on various options for producing hydrogen from Victorian brown coal.[4] The purpose of this work was to explore the pathways of hydrogen production from Victorian brown coal through a thermal process. The target hydrogen production rate was set at 770 tonnes/day, as an estimate of the quantity of hydrogen exported from Victoria to Japan in the 2030s.

The four options explored were:

- Brown coal pyrolysis plant with tar and carbon dioxide removal by calcium looping for hydrogen production by pressure swing adsorption
- Brown coal pyrolysis plant followed by tar removal, reforming, shift reactors and carbon dioxide removal by calcium looping for hydrogen production by pressure swing adsorption
- Brown coal pyrolysis plant followed by tar removal, reforming, shift, tar cracking reactors and carbon dioxide removal by calcium looping for hydrogen production by pressure swing adsorption
- Brown coal gasification plant using oxygen blown entrained for gasifier followed by shift reactor for hydrogen production by pressure swing adsorption.

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[2] COAG Energy Council (2018). *Establishment of the Hydrogen Working Group of the COAG Energy Council* http://www.coagenergycouncil.gov.au/publications/establishment-hydrogen-working-group-coag-energy-council [Accessed Dec. 2018]

[3] Constable, T. (2019). *Using Australian coal to power hydrogen's future*. Minerals Council of Australia, Canberra, Australia. https://www.minerals.org.au/news/using-australian-coal-power-hydrogen's-future [Accessed Jan 2019] – reproduced in full in the Appendix.

[4] Kibria, M.A. and Bhattacharya, S. (2018). *Process Simulation of Hydrogen Production Options from Victorian Brown Coal.* Monash University, Clayton, Victoria

Monash University and CSIRO used a process simulation software to model the various processes. This current report seeks to provide a both capital cost and a levelized cost of hydrogen for those options.

11 Cost Methodology

Typically, as a project evolves, it becomes more definitive, just as cost estimates of evolving projects also become more definitive over time. Given the early stage of this project, a Class 5 cost estimate has been prepared, with some elements of the estimate Class 4 as defined by the Association for Advancement of Cost Engineering International (AACEI) is summarised in Table 12.[5]

Table 12 Cost estimate classifications and primary characteristics

Cost Estimate Classification	**Primary Characteristics**	
	Level of Definition (% of completed definition)	Cost Estimating Description (techniques)
Class 5 Concept Screening	0 – 2	Stochastic, most parametric, judgement (parametric, specific analogy, expert opinion, trend analysis)
Class 4 Study or Feasibility	1 to 15	Various, more parametric, (parametric, specific analogy, expert opinion, trend analysis)
Class 3 Preliminary, Budget Authorisation	10 to 40	Various, including combinations (detailed, unit-cost or activity-based, parametric, specific analogy, expert opinion, trend analysis)
Class 2 Control or Bid/Tender	30 to 70	Various, more definitive (detailed, unit-cost or activity-based, expert opinion, learning curve)
Class 1 Check Estimate or Bid/Tender	50 to 100	Deterministic, most definitive (detailed, unit-cost or activity-based, expert opinion, learning curve)

The analysis has not attempted to quantify all the uncertainties or ranges in a rigorous manner due to the estimate Class. However, it should be noted that some costs have greater uncertainty than others. There is also some uncertainty due to the process configuration due to the early stage modelling that has been completed.[5] Due to a lack of similar or fully analogous Australian projects with known cost data, some data has been sourced from 'older' data or factored from comparable data sets.

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[5] Sanchez, R. (2011). *Cost Estimating Guide*. DOE G 413.3-21

11.1 Basis of Costs

The cost estimates have been prepared as Total Overnight Costs in 2019 AUD, covering bare erected costs engineering, procurement, construction management, as well as owners' costs and process and project contingencies in accordance with the AACE recommendations for the estimate class.

Bare Erected Cost (BEC) comprises the delivered cost of process equipment, bulk materials, on-site facilities and infrastructure that support the plant and the direct and indirect labour, equipment and consumables required for its construction and/or installation. The cost of EPC services and contingencies are not included in BEC.

Engineering, Procurement, and Construction Cost (EPCC) comprises the BEC plus the cost of services provided by the engineering, procurement, and construction (EPC) contractor. EPC services include detailed design, procurement, contractor permitting, and project/construction management costs.

Total Plant Cost (TPC) comprises the BEC, EPCC plus project and process contingencies.

Total Overnight Capital (TOC) comprises the TPC plus all other overnight costs, including owner's costs and any allowances that have been applied to identified additional items not yet quantified. TOC is an "overnight" cost, expressed in base-year dollars and as such does not include escalation or interest during construction.

Total As-Spent Capital (TASC) is the sum of all capital expenditures as they are incurred during the capital expenditure period including their escalation. TASC also includes interest during construction. Accordingly, TASC is expressed in mixed, current-year dollars over the capital expenditure period.

The various cost descriptions are shown in *Figure 7*.[6]

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[6] National Energy Technology Laboratory (2011). *Quality guidelines for energy system studies. Cost Estimation methodology for NETL assessments of power plant performance*. DOE/NETL-2011/1455

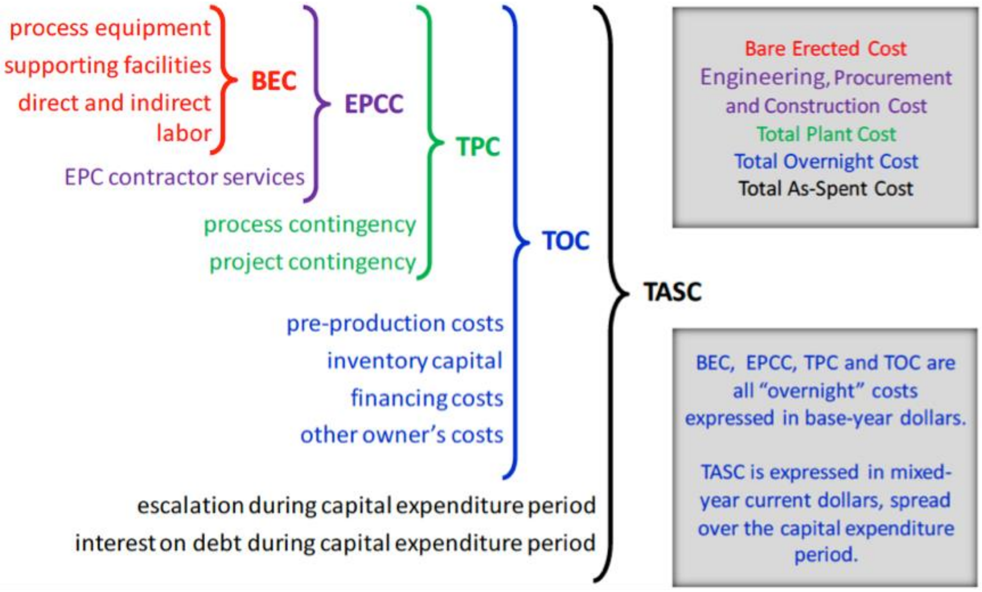

Figure 7 Capital cost methodology – breakdown of different cost levels

The capital costs for this study were prepared as of the 1st of January 2019 in AUD, and then adjusted to 2030 by using the Reserve Bank of Australia midpoint estimate for inflation.[7] Learn by doing cost reductions were applied to the most critical processes before escalations, using the approach adopted in CSIRO's National Hydrogen Roadmap study.

The 2019 cost estimates were based on both public and confidential information (available to the authors of this report). A confidential peer review of the data was also conducted, and recommended adjustments were taken into account.

11.2 Levelised Cost of Product

The financial model inputs are summarised *Table 13* and input costs in *Table 14*.

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[7] PricewaterhouseCoopers Consulting (2017). *Cost escalation factors. Final report*. www.pwc.com.au. [Accessed Dec. 2018]

Table 13 Base case financial assumptions

Parameter	Unit	Value
Cost of Equity	%	10
ROI Treasuries	%	4
ROI Market	%	9
Volatility	- -	1.2
Cost of Debt	%	5
90-day Bank Rate	%	4
Risk Premium	%	1
WACC	%	7
Debt	%	60
Equity	%	40
Length of loan	Years	10
Tax Rate	%	30
Economic Life	Years	30
Operating Hours in Year	Hours	7074*

* based on 85% availability and 95% utilisation

Table 14 Input costs assumptions

Inputs	Unit	Base Case
Brown Coal	$/tonne	10
Char	$/tonne	450
CO_2 Sequestration[8]	$/tonne	10
Electricity	c/kWh	7

The Levelised Cost of a Product (LCoP) seeks to account for all physical assets and resources required to deliver one unit of the product. LCoP could be defined (similar to Levelised Cost of Electricity or LCOE) as the constant dollar product price that would be required over the life of the plant to cover all operating expenses, payment of debt and accrued interest on initial project expenses, and the payment of an acceptable return to investors. Per unit of output, the LCoP

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[8] The cost of CO_2 sequestration is based on an estimate of $9.30/t CO_2 injected, provided in CO2CRC's 'Australian Power Generation Technology' report (2015), Fig. 140, for the case of 5 million t/y CO_2 being injected under the nearshore Gippsland Basin (http://www.co2crc.com.au/wp-content/uploads/2016/04/LCOE_Report_final_web.pdf)

aggregates a share of the initial capacity investment with periodic fixed and variable operating costs.

A Discounted Cashflow (DCF) methodology using the parameters listed in *Table 13* and *Table 14* was used to calculate the LCoP of hydrogen on the following basis:

- Capital costs where factored from relevant equipment costs for each option.
- Capital flows included interest and loan repayments for 60% debt funding as well as sustaining capital based on a 5 year turn around frequency.
- Capital costs included allowances for design growth, yard piping and piperacks, first fills and commissioning costs. Since Option 1 resulted in the production of a large quantity of carbon rich PSA tailgas a further allowance was made for its re-processing.
- Income was made up from sale of product at cost. No income from any surplus power was included since there was insufficient information regarding parasitic loads as well as energy flows from the normally expected combustion of PSA tail gas and un-processed tar oils.
- A Straight-Line Depreciation method based on a 5% per annum deprivation rate was used with an initial accelerated depreciation rate of 20% over the first 5 years.
- Expenses were divided into fixed and variable and covered raw materials, catalysts and chemicals, CO_2 sequestration costs, operations and maintenance labour, maintenance materials, management labour, marketing costs, insurance and external services.

12 Literature Overview

A literature survey was conducted to examine the recent and relevant publicly available information on hydrogen production from fossil fuels. The reports and websites that have been reviewed are listed in *Table 19*, and a summary of each resource is covered in the following section.

In summary, the level of detail and currency of fossil fuel-based hydrogen production process is poor. The most detailed references are based on pre-2010 data, which is difficult to covert to the current Australian market conditions. Additionally, the major focus for most research and development institutions is on ‘green’ renewable based hydrogen production, with little to no comparisons with non-renewable alternatives. The most relevant Australian reference is CSIRO’s recent Hydrogen Roadmap[9] published in 2018, containing comparative fossil fuel and renewable hydrogen generation costs, along with ‘best’ case scenarios looking at technology improvements. Unfortunately, while up-to-date, the cost and performance data are relatively high level and does not cover any pyrolysis opportunities which would be relevant for this study.

12.1 Summary of Hydrogen Production Resources

The International Energy Agency Hydrogen Technology Collaboration Program (IEA-TCP)[10] is an international coordinated hydrogen research, development and demonstration program. The basic “unit of organization” is the Task/Annex, a research project that focus on a particular facet of hydrogen and typically they are allotted three years to be completed. The dominant focus of the work is listed in Table 15, and the IEA TCP focus more broadly is on renewable based hydrogen, given that “renewables relationship with hydrogen cannot be overemphasised.”[11] Unfortunately for this current cost report, the fossil fuel-based work is also quite dated, having been completed as Task 16 in 2005 (refer to Appendix B: IEA Hydrogen Activity List) Being more than 10 years old, the costs have not been reviewed. The IEA TCP website10 however, is a comprehensive and transparent resource on Hydrogen, and its recent report on the global trends and outlook 11 is a useful reference in this regard.

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[9] Bruce, S., Temminghoff, M., Hayward, J., Schmidt, E. Munnings, C., Palfreyman, D. and Hartley, P. (2018). *National Hydrogen Roadmap.* CSIRO, Australia

[10] ieahydrogen.org. (2018). *Hydrogen Implementing Agreement*. [online] http://ieahydrogen.org/Home.aspx [Accessed Nov. 2018]

[11] De Valladares, M-R. (2017). *Global Trends and Outlook for Hydrogen.* IEA Hydrogen Technology Collaboration Program

Table 15 IEA-TCP current task list (with hyperlinks)

Current Tasks		
Task 32	H_2 Based Energy Storage	2013-2016
Task 33	Local H_2 Supply for Energy Applications	2013-2016
Task 34	Biological Hydrogen for Energy and Environment	2014-2017
Task 35	Renewable Hydrogen Production	2014-2017
Task 36	Life Cycle Sustainability Assessment	2014-2017
Task 37	Hydrogen Safety	2015-2018
Task 38	Power to Hydrogen	-
Task 39	Hydrogen in Marine Applications	-

The National Energy Technology Laboratory's website contains a Technologies for Hydrogen Production page which highlights the key fossil fuel technologies and an extensive technology efficiency and cost comparison table. Unfortunately, these are based on a 2004 reference and as such are not useful for this current study.[12] The site is fully integrated with other elements of NETL's technology assessments, much of which contains relevant and up-to-date information on how hydrogen may be produced.

The Energy Efficiency & Renewable Energy website[13] of the DOE details hydrogen production R&D from mainly renewable energy sources. The site also covers a range of hydrogen related issues, including legal and regulatory. While the DOE does have some references to coal related hydrogen production[14] (which is not easily accessible from its landing pages), it does not have any cost information.

The Comparison of Dispatchable Renewable Electricity Options report is a renewables-based report for the Australian Renewable Energy Agency (ARENA) with a strong emphasis on electricity generation. It does contain some limited hydrogen costing information, which is the most recent public data in the Australian. While it does not aim to cover any fossil fuel-based hydrogen options, one hydrogen option covered is relevant (summarised in *Table 16*).[15]

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[12] T-Raissi, A. and Block, D. (2004). *Hydrogen: Automotive Fuel of the Future.* IEEE Power & Energy, Vol. 2, No. 6, page 43, Nov-Dec 2004

[13] afdc.energy.gov. (2018). *Alternative Fuels Data Center: Hydrogen Research and Development.* [online] https://afdc.energy.gov/fuels/hydrogen_research.html [Accessed Nov. 2018]

[14] energy.gov. (2018). *Hydrogen Production: Coal Gasification.* [online] https://www.energy.gov/eere/fuelcells/hydrogen-production-coal-gasification [Accessed Nov. 2018]

[15] Lovegrove, K., James, G., Leitch, D., Milczarek,. Ngo, A., Rutovitz, J, Watt, M. and Wyder, J. (2018). *Comparison of Dispatchable Renewable Electricity Options.* ITP Thermal P/L.

Table 16 Electrolyser based hydrogen cost data

	Installed cost	Baseline capacity	Power law size exponent
Electrolyser	\$1.09m/MW$_e$	20 MW$_e$	0.7
Plus underground storage	\$0.66m/MWh$_t$	20,000 MWh$_t$	0.7
Plus combined cycle gas turbine	\$1.64m/MW$_e$	20 MW$_e$	0.7

In 2018, CSIRO released the Australian National Hydrogen Roadmap[9], which explored pathways to an economically sustainable hydrogen industry in Australia. Its objective was to provide a blueprint for the development of a hydrogen industry in Australia, building on current projects and to help inform the next series of investment amongst industry, government and research institutions so that the industry can continue to scale in a coordinated manner. The Roadmap concludes that hydrogen could be produced from brown coal at approximately \$2.14 -2.74/kg once the commercial scale production and CCS plant come online in the 2030s.[16] The brown coal-based hydrogen production cost and technology performance is summarized in *Table 17*.[9]

Table 17 Best case data for brown coal gasification with CCS

Key Cost Driver	**Unit**	**'Best Case'**
Coal price	\$/GJ	1.5
Utilisation factor	%	85
Scale/Capacity	t H_2/day	770
Asset life	Years	40
Capex (less risk)	\$ kg H_2/day	10.45
Fixed opex	\$ kg H_2/day	0.41
Variable opex	\$ kg H_2/day	0.04
Cost of CO_2 storage	\$/t CO_{2-e}	7 - 40
Efficiency	%	67
Other revenue	\$m year	25.5
Risk	%	15
Real discount rate	%	7
Cost of capital	%	5
Levelised cost of H_2	**\$/kg**	**2.14 – 2.74**

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[16] The modelling that underpins these figures was undertaken by CSIRO and only contains information that has been made publicly available by members of the Hydrogen Energy Supply Chain project.

The IEA's recent report on Global Trends and Outlook for Hydrogen,[11] realistically sets the scene for the potential roles for hydrogen usage in the context of deep global carbon abatement. The report is a high-level document, while giving cost ranges for some applications, they are provided with little to no context, meaning the data is of little use to this study. The reference to carbon capture and storage (CCS) acknowledges a likely use of the technology at scale, however contains no detail on how this will occur, or any costs associated with this approach.

A review of the Uniper Energy Storage Power-To-Gas (P2G) demonstration projects at Falkenhagen and Hamberg-Reitbrook in Germany was conducted for the Electric Power Research Institute (EPRI).[17] This report details the power to hydrogen work using grid sourced power, with a view to demonstrating the concept, with a bias to sourcing renewable energy to power the technology in the future. While it contains no references or links to fossil fuel-based hydrogen production, it is a quality reference on the production of hydrogen in the European context. Power to gas is a competitor to the technologies[4] modelled for this current work.

CSIRO's Low Emissions Technology Roadmap[18] seeks to identify the key commercial opportunities for industry that low emissions technologies in the technology sector can provide. In "Pathways 3 and 4" the role hydrogen is investigated as an energy storage medium across the energy sector. As part of this analysis, gasification linked CCS has been identified as enabler of local 'low emissions' hydrogen production, and a key export opportunity. While hydrogen 'levelised costs of production' are reported within the Roadmap executive summary and technical report, the report does not contain cost information that can be used in this this current work. According to the Roadmap, low or zero emissions hydrogen is most likely to be produced at large-scale using:

- Electrolysis using dedicated renewables: ~$8-10/kg in 2030 with zero emissions
- Coal gasification with CCS: ~$2-3/kg in 2030 with higher emissions intensity than electrolysis (7-15 kg CO_{2e}/kg H_2)[18]

Chemical looping was included in this literature review as part of costing the capture technology chosen by Monash University and CSIRO.[4] Chemical looping in this application has a low technical readiness level, and as such cost and performance data is at the low end of costing accuracy (refer also *Table 12*). The paper by Mantripgragada and Rubin[19] sought to compare cost and performance of chemical looping with the more conventional Selexol. While they found the chemical looping performed better in terms of overall energy penalties, it was more capital intensive.

A comparative overview of the major hydrogen production methods was carried out.[20] This included detailing clear and transparent process descriptions along with the technical aspects of different methodologies covered. The overall conclusions show that thermo-chemical pyrolysis

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[17] Electric Power Research Institute (EPRI), (2017). *Program on Technology Innovation: Review of the Uniper Energy Storage GmbH Power-To-Gas (P2G) Demonstration Projects at Falkenhagen and Hamberg-Reitbrook, Germany.* EPRI, Palo Alto, CA: 3002011519

[18] Campey, T., Bruce, S., Yankos, T., Hayward, J., Graham, P., Reedman, L., Brinsmead, T., Deverell, J. (2017). *Low Emissions Technology Roadmap.* CSIRO, Australia. Report No. EP167885

[19] Mantripragada, H. and Rubin, E. (2017). *Chemical Looping for Pre-combustion and Post-combustion CO_2 Capture.* Energy Procedia, 114, pp.6403-6410

[20] Nikolaidis, P. and Poullikkas, A. (2017). *A comparative overview of hydrogen production processes.* Renewable and Sustainable Energy Reviews, 67, pp.597–611

and gasification are the currently economically viable approaches providing the highest potential to become competitive on a large scale in the near future. These conventional methods have an estimated hydrogen production with costs in the range of $USD 1.34 – 2.27 per kg. Unfortunately, while this review was published in 2017, its costs for most of the technologies are based on references ranging from 2002 and 2007, with one reference from 1992 and solar/biomass related technologies from 2014. This makes the results and costings from this study not useful as an input into this current work.

While not a report that covers hydrogen production technologies, the Five Keys to Unlock CCS Investment[21] by the IEA, it discusses CCS is a key enabling technology for low emissions hydrogen from fossil fuels. The IEA has consistently highlighted the critical role of CCS in meeting global climate objectives, and its "five keys to unlock CCS investment" are:

1. Harvest "low-hanging fruit" to build CCS deployment and experience from the ground up.
    - The production of hydrogen is highlighted as a 'low hanging fruit' option.
2. Tailor policies to shepherd CCS through the early deployment phase and to address the unique integration challenges for these facilities.
3. Target multiple pathways to reduce costs from technological innovation in carbon capture and CO_2 utilisation to progressive financing arrangements.
4. Build CO_2 networks and accelerate CO_2 storage assessments in key regions.
5. Strengthen partnerships and co-operation between industry and governments.

Without a viable CCS options, there is little scope for low emissions hydrogen from fossil fuels.

An investigation into the supply chain feasibility carried out by Kawasaki Heavy Industries (KHI) on brown coal CO_2-free hydrogen from Australia to Japan.[22] Within this review, high level supply costs are noted, with a percentage breakdown given for the various components of the supply chain (refer to *Figure 8*) at a total cost of ¥43.6bn. However, these costs are not very transparent or clear in terms of what is included. In addition, no financial parameters are provided for calculating levelized cost of production.

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[21] Carboncapturejournal.com. (2017). *IEA holds global summit on carbon capture.* [online] Available at: http://www.carboncapturejournal.com/ViewNews.aspx?NewsID=3977 [Accessed Nov. 2018]

[22] Kamiya, S., Nishimura, M. and Harada, E. (2018). *Study on Introduction of CO_2 Free Energy to Japan with Liquid Hydrogen.* Physics Procedia, 67(2015), pp.11-19

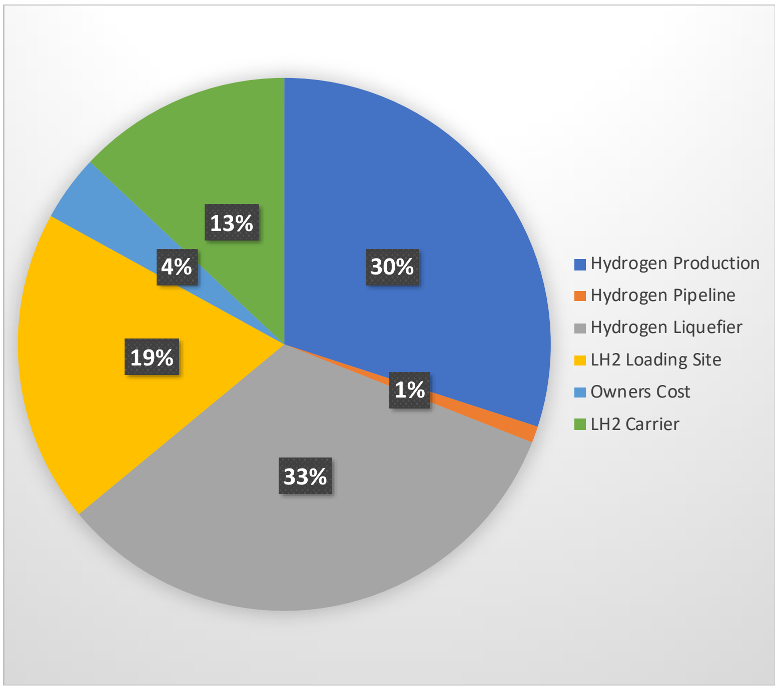

Figure 8 Cost breakdown of overall hydrogen production facilities

The Assessment of Hydrogen Production with CO_2 Capture, Volume 1: Baseline State-of-the-Art Plants,[23] is part of a series of reports that the US Department of Energy regularly sponsor. These reports are considered important benchmarks. This report is a detailed piece of work based on the National Energy Technology Laboratory (NETL) preparing plant designs which are then costed using verifiable sources. The report focuses on steam reforming of natural gas and coal gasification with and without CCS. While the report was published in 2010, the cost basis is 2007, and as such, the costs should be used with caution (refer to *Table 18*). The report does not compare the fossil fuel costs reported with any other technologies.

Table 18 Capital cost estimation results (June 2007 USD)

Case	Unit	Steam reforming of natural gas	Coal gasification with CCS
H_2 Production	(kg H_2/day)	618,936	618,940
Bare Erected Cost	($'000s)	$343,355	$958,576
Eng, CM, HO, Fees, etc.	($'000s)	$34,335	$95,858
Project Contingency	($'000s)	$83,107	$190,503
Process Contingency	($'000s)	$32,909	$61,446
Total Plant Cost	($'000s)	**$493,706**	**$1,306,383**
Total Plant Cost	$/ (kg H_2/day)	**$798**	**$2,111**

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[23] National Energy Technology Laboratory (NETL) – US Department of Energy (DOE), (2010). *Assessment of Hydrogen Production with CO_2 Capture. Volume 1: Baseline State-of-the-Art Plants.* DOE/NETL-2010/1434

Owner's Cost	(\$'000s)	\$118,926	\$291,118
Total Overnight Cost	(\$'000s)	**\$612,632**	**\$1,597,501**
Total Overnight *Cost*	\$USD/ (kg H_2/day)	**\$990**	**\$2,581**

Table 19 List of key reports and data sources.

Name	**Organisation**	**Date**
IEA Hydrogen website[10]	International Energy Agency	Current
National Energy Technology Laboratory website[24]	National Energy Technology Laboratory	Current
Energy Efficiency & Renewable Energy Website[13]	US Department of Energy: Energy Efficiency & Renewable Energy	Current
Comparison of Dispatchable Renewable Electricity Options[15]	ITP Energised Group, Institute for Sustainable Futures, ITK Consulting – for ARENA	June 2018
National Hydrogen Roadmap[9]	CSIRO	2018
Global Trends and Outlook for Hydrogen[11]	International Energy Agency: Hydrogen Technology Collaboration Project	December 2017
Program on Technology Innovation: Review of the Uniper Energy Storage GmbH Power-To-Gas (P2G) Demonstration Projects at Falkenhagen and Hamberg-Reitbrook, Germany[17]	Electric Power Research Institute	August 2017
Low Emissions Technology Roadmap[18]	CSIRO	June 2017
Chemical Looping for Pre-Combustion and Post-Combustion CO_2 Capture[19]	Carnegie Mellon University	2017
A Comparative Overview of Hydrogen Production Processes[20]	Department of Electrical Engineering, Cyprus University of Technology	2017
Five Keys to Unlock CCS Investment[21]	International Energy Agency	2016
Study on Introduction of CO_2 free Energy to Japan with Liquid Hydrogen[22]	Kawasaki Heavy Industries Ltd	2015
Assessment of Hydrogen Production with CO_2 capture. Volume 1: Baseline State-of-the-Art Plants[23]	National Energy Technology Laboratory – U.S. Department of Energy	August 2010.

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[24] netl.doe.gov. (2018). *Hydrogen*. [online] https://www.netl.doe.gov/research/coal/energy-systems/gasification/gasifipedia/technologies-hydrogen [Accessed Nov. 2018]

13 Option Costing

13.1 Option 1: Pyrolysis plant

The raw Victorian brown coal is processed by subsequent two-step hammering and screen to the desired particle size for efficient drying. The raw coal loses almost 55% of its initial weight during the drying process. The dry coal then goes to a briquette unit in which the dry coal is compressed to cylindrical pellets. These pellets are then fed to a pyrolyser unit. A certain amount of char is combusted to produce the heat energy to run the pyrolyser and to produce steam for the steam tube dryer. The flue gas is then quenched at low temperature to separate the tar and subsequently passed to a calcium carbonate looping system for CO_2 separation. The CO_2 free gases then moved through a pressure swing adsorption process to separate the hydrogen. The simplified block flow diagram is shown in Figure 9.

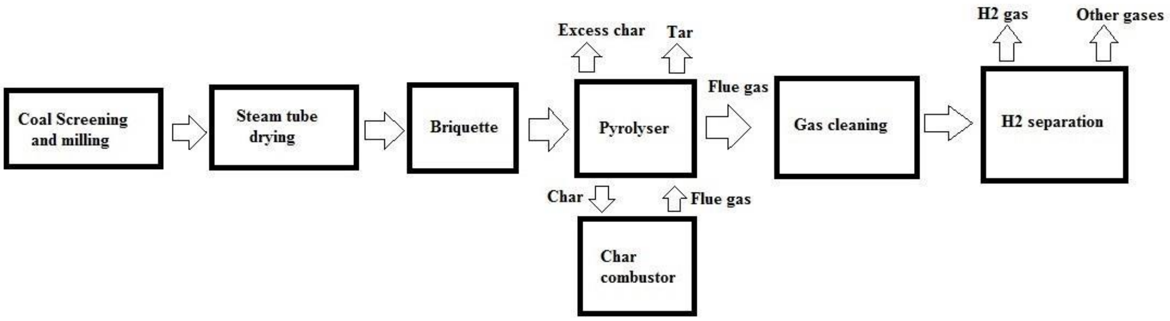

Figure 9 Block diagram of Option 1

The key parameters and results obtained from the study are summarised in *Table 20* and *Table 21* respectively. The corresponding sensitivity cost analysis is shown in *Table 22*.

Table 20 Option 1 – Key high-level parameters.

Item	**Value (tonne/h)**
Target hydrogen	32.1
Raw coal	3680
Dry coal	1136
Excess char to refinery	69
Tar production	124
Total carbon dioxide generated	1391
Steam requirement	173

Table 21 Option 1 – Capital costs, 2019 AUD

Area	Description	2019
1	Coal Screening and Milling	587,000
2	Steam Tube Drying	651,000
3	Secondary Screening and Milling	508,000
4	Briquetting	1,155,000
5	Pyrolysis	316,000
6	Char Combustion / Boiler	537,000
7	Tar Separation	256,000
8	CO_2 Removal	3,239,000
9	CO_2 Compression and Dehydration	336,000
10	H_2 Separation	3,350,000
11	Power Generation	-
12	Balance of Plant	1,091,000
13	Infrastructure	420,000
14	Indirect Costs	524,000
15	Allowances	3,435,000
16	Contingency	3,891,000
	TOTAL	**$20,297,000**

Table 22 Option 1 – Sensitivity studies: 2019

			Variable							
			CAPEX		**Char**		**CO_2 Seq.**		**WACC**	
Sensitivity	**Units**	*Base Case*	High	Low	High	Low	High	Low	High	Low
CAPEX	$ Mil	20,297	28,416	15,223	20,297	20,297	20,297	20,297	20,297	20,297
Coal feed (AR)	tonne/hr	3,680	3,680	3,680	3,680	3,680	3,680	3,680	3,680	3,680
Char	tonne/hr	69	69	69	69	69	69	69	69	69
CO_2 produced	MMt/y	15	15	15	15	15	15	15	15	15
Char price	$/tonne	450	450	450	600	300	450	450	450	450
WACC	%	7	7	7	7	7	7	7	10	5
CO_2 Seq.	$/tonne	10	10	10	10	10	15	5	10	10
Levelised cost of H_2	**$/kg**	**11.39**	**15.58**	**8.77**	**11.07**	**11.72**	**11.61**	**11.17**	**13.80**	**9.90**

13.2 Option 2: Pyrolysis plant followed by shift reactor

Option 2 is similar to Option 1, except it includes a water gas shift unit coupled with methane steam reformer unit. The flue gas leaving the pyrolyser unit is rich in carbon monoxide and methane. To increase the hydrogen production level, the available methane is transformed to hydrogen and carbon monoxide in the methane steam reformer unit. The simplified block flow diagram is shown in Figure 10.

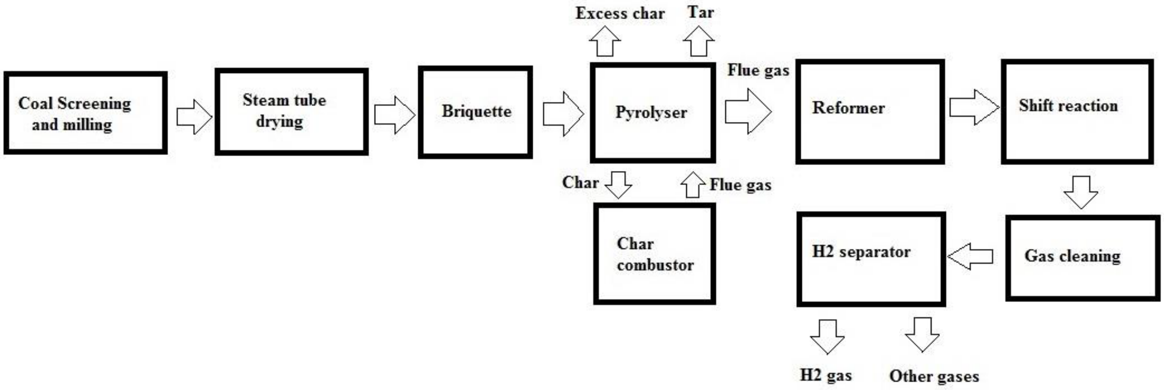

Figure 10 Block diagram of Option 2

The key parameters and results obtained from the study are summarised in *Table 23* and *Table 24* respectively. The corresponding sensitivity cost analysis is shown in *Table 25*.

Table 23 Option 2 – Key high-level parameters.

Item	**Value (tonne/h)**
Target hydrogen	32.1
Raw coal	1780
Dry coal	550
Excess char to refinery	83
Tar production	62
Total carbon dioxide generated	924
Steam requirement	375

Table 24 Option 2 – Capital costs, 2019 AUD.

Area	Description	2019
1	Coal Screening and Milling	329,000
2	Steam Tube Drying	338,000
3	Secondary Screening and Milling	247,000
4	Briquetting	563,000
5	Pyrolysis	154,000
6	Char Combustion / Boiler	214,000
7	Tar Separation	98,000
8	Reforming	777,000
9	Shifting	160,000
10	CO_2 Removal	1,499,000
11	CO_2 Compression and Dehydration	228,000
12	H_2 Separation	1,457,000
13	Power Generation	-
14	Balance of Plant	706,000
15	Infrastructure	213,000
16	Indirect Costs	373,000
17	Allowances	794,000
18	Contingency	2,207,000
	TOTAL	**$10,395,000**

Table 25 Option 2 – Sensitivity studies: 2019.

			Variable							
			CAPEX		**Char**		**CO_2 Seq.**		**WACC**	
Sensitivity	**Units**	*Base Case*	High	Low	High	Low	High	Low	High	Low
CAPEX	$ Mil	10,359	14,503	7,769	10,359	10,359	10,359	10,359	10,359	10,359
Coal feed (AR)	tonne/hr	1,780	1,780	1,780	1,780	1,780	1,780	1,780	1,780	1,780
Char	tonne/hr	83	83	83	83	83	83	83	83	83
CO_2 produced	MMt/y	870	870	870	870	870	870	870	870	870
Char price	$/tonne	450	450	450	600	300	450	450	450	450
WACC	%	7	7	7	7	7	7	7	10	5
CO_2 Seq.	$/ton	10	10	10	10	10	15	5	10	10
Levelised cost of H_2	**$/kg**	**5.15**	**7.29**	**3.78**	**4.76**	**5.54**	**5.29**	**5.01**	**6.38**	**4.38**

13.3 Option 3: Pyrolysis plant followed by shift and tar cracking reactor

Option 3 is similar to Option 2, except this option has chemical looping-based cracking process. The quenched and cooled flue gas after pyrolysis has a significant amount of tar, which goes in a tar cracking unit. In this unit, the long chain hydrocarbons are broken down to light hydrocarbons and passed through the methane steam reforming and water gas shift reactor. The simplified block flow diagram is shown in Figure 11.

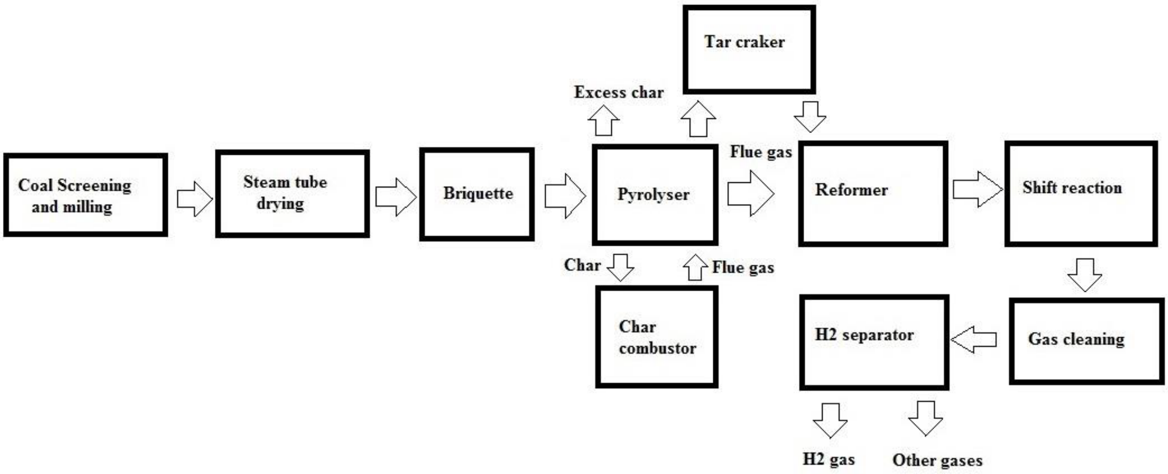

Figure 11 Block diagram of Option 3

The key parameters and results obtained from the study are summarised in *Table 26* and *Table 27* respectively. The corresponding sensitivity cost analysis is shown in *Table 28*.

Table 26 Option 3 – Key high-level parameters.

Item	**Value (tonne/h)**
Target hydrogen	32.1
Raw coal	1480
Dry coal	457
Excess char to refinery	67
Tar production	0
Total carbon dioxide generated	902
Steam requirement	340

Table 27 Option 3 – Capital costs, 2019 AUD.

Area	**Description**	**2019**
1	Coal Screening and Milling	283,000
2	Steam Tube Drying	292,000
3	Secondary Screening and Milling	214,000
4	Briquetting	485,000
5	Pyrolysis	135,000
6	Char Combustion / Boiler	183,000
7	Tar Separation	88,000
8	Tar Cracking	119,000
9	Reforming	729,000
10	Shifting	150,000
11	CO_2 Removal	1,400,000
12	CO_2 Compression and Dehydration	225,000
13	H_2 Separation	1,154,000
14	Power Generation	-
15	Balance of Plant	610,000
16	Infrastructure	186,000
17	Indirect Costs	332,000
18	Allowances	641,000
19	Contingency	1,976,000
	TOTAL	**$9,205,000**

Table 28 Option 3 – Sensitivity studies: 2019.

			Variable							
			CAPEX		Char		CO_2 Seq.		WACC	
Sensitivity	**Units**	*Base Case*	High	Low	High	Low	High	Low	High	Low
CAPEX	$ Mil	9,205	12,888	6,904	9,205	9,205	9,205	9,205	9,205	9,205
Coal feed (AR)	tonne/hr	1,480	1,480	1,480	1,480	1,480	1,480	1,480	1,480	1,480
Char	tonne/hr	67	67	67	67	67	67	67	67	67
CO_2 produced	MMt/y	854	854	854	854	854	854	854	854	854
Char price	$/tonne	450	450	450	600	300	450	450	450	450
WACC	%	7	7	7	7	7	7	7	10	5
CO_2 Seq.	$/tonne	10	10	10	10	10	15	5	10	10
Levelised cost of H_2	**$/kg**	**4.67**	**6.58**	**3.46**	**4.36**	**4.99**	**4.81**	**4.50**	**5.77**	**4.00**

13.4 Option 4: Oxy-blown gasification followed by shift reactor

The raw Victorian brown coal is processed by subsequent two-step hammering, screen and drying process to the desired particle size before feeding to the entrained flow gasifier. The entrained flow gasifier used in the simulation is a two-stage oxy-fired slagging gasifier which mimics the Mitsubishi Heavy Industry (MHI) gasifier. The syngas after gasifier unit is carbon monoxide rich. The syngas is cooled and subsequently passed through a shift reactor to increase the hydrogen level. The syngas goes to a chemical looping system for carbon dioxide separation. The carbon dioxide free gases then passed through a pressure swing adsorption process to separate the hydrogen. The simplified block flow diagram is shown in Figure 12.

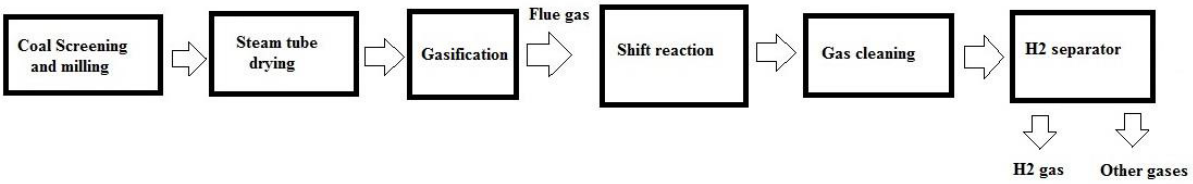

Figure 12 Block diagram of Option 4

The key parameters and results obtained from the study are summarised in *Table 29* and *Table 30* respectively. The corresponding sensitivity cost analysis is shown in *Table 31*.

Table 29 Option 4 – Key high-level parameters.

Item	**Value (tonne/h)**
Target hydrogen	32.1
Raw coal	893
Dry coal	332
Excess char to refinery	0
Tar production	0
Total carbon dioxide generated	518
Steam requirement	344

Table 30 Option 4 – Capital costs, 2019 AUD

Area	Description	2019
1	Coal Screening and Milling	143,000
2	Steam Tube Drying	182,000
3	Secondary Screening and Milling	144,000
4	Air Separation	447,000
5	Gasification	1,106,000
6	Shift	537,000
7	CO_2 Removal	489,000
8	CO_2 Compression and Dehydration	98,000
9	H_2 Separation	264,000
10	Power Generation	179,000
11	Balance of Plant	446,000
12	Infrastructure	186,000
13	Indirect Costs	288,000
14	Allowances	573,000
15	Contingency	1,216,000
	TOTAL	**$5,844,000**

Table 31 Option 4 – Sensitivity studies: 2019

			Variable					
			CAPEX		**CO_2 Seq.**		**WACC**	
Sensitivity	**Units**	*Base Case*	High	Low	High	Low	High	Low
CAPEX	$ Mil	5,844	8,182	4,383	5,844	5,844	5,844	5,844
Coal feed (AR)	tonne/hr	893	893	893	893	893	893	893
CO_2 produced	MMt/y	491	491	491	491	491	491	491
WACC	%	7	7	7	7	7	10	5
CO_2 Seq.	$/tonne	10	10	10	15	5	10	10
Levelised cost of H_2	**$/kg**	**3.51**	**4.71**	**2.77**	**3.59**	**3.43**	**4.23**	**3.09**

13.5 Cost Summary - Options 1 to 4

The costs for all four options where originally calculated in 2019 AUD. A 'learn by doing' factor was then applied as relevant, and then the costs escalated to 2030. These and the related levelized cost of hydrogen for the base cases are listed in *Table 32*.

Table 32 Base case cost summaries: 2019, 2030 & LCoP

Option	**2019**		**2030**	
	CAPEX ($M)	LCoP ($/kg)	CAPEX ($M)	LCoP ($/kg)
1	20,297	11.39	22,104	12.33
2	10,359	5.15	11,281	5.62
3	9,205	4.67	10,025	5.10
4	5,844	3.51	5,751	3.46

Options 1 to 3 are based on briquette production and pyrolysis to manufacture hydrogen. As such, less learn by doing is anticipated compared with Option 4, as there is likely to be industry wide learning within the gasifier and gas processing associated with hydrogen production. The results in Options 1 to 3 increasing in capital costs by 2030, while Option 4 shows a small decrease. Realistically, all options change little between now and 2030 based on the learn-by-doing and inflation assumptions given the accuracy of these estimates.

As shown in *Table 22*, *Table 25*, *Table 28* and*Table 31*, the LCoP is most sensitive to variations in the estimated CAPEX requirements. *Table 33* below shows the likely range of 2019 and 2030 LCoP values, based on their sensitivity to CAPEX estimates.

Table 33 CAPEX sensitivity analysis cost summaries: 2019, 2030 & LCoP

Option	**2019**		**2030**	
	CAPEX ($M)	LCoP ($/kg)	CAPEX ($M)	LCoP ($/kg)
1	15,223 – 28,416	8.77 – 15.58	16,578 – 30,946	9.49 – 16.86
2	7,769 – 14,503	3.78 – 7.29	8,460 – 15,794	4.47 – 8.61
3	6,904 – 12,888	3.46 – 6.58	7,519 – 14,036	3.78 – 7.19
4	4,383 – 8,182	2.77 – 4.71	4,318 – 8,052	2.73 – 4.64

13.6 Study Observations

This section of the report covers some of the authors' observations and comments on the base study, in addition to comments on what has been adapted to ensure a comprehensive cost and comparison basis could be achieved.

The PSA tail gas in Option 1 has been reprocessed through the plant in order to remove its carbon content.

Power generation equipment costs have not been incorporated to Option 1, 2 or 3, as per the Monash-CSIRO modelling. It is likely in the next phase of study that an options analysis for heat integration and power generation be compared with purchased power.

Option 1 does not reform and shift its tar free gas and consequently does not capture all the carbon in the pyrolysis synthesis gas. Whilst perhaps outside the scope of this work, the Study Team has estimated that combusting the additional carbon monoxide and methane would increase the CO_2 content of Option 1 by approximately 50%. Cracking the tar oils as in Option 3 or simply combusting these would further increase the produced CO_2; however, this would be expected to be relatively small in comparison.

The implications of not including equivalent CO_2 capture and compression on all options disadvantages the lowest CO_2 producing option, and favours the high CO_2 producing option, as CO_2 capture and compression are both energy intensive. Both, especially capture, also have significant capital impacts.

The Option 1, Combustor and Heat Recovery Unit in the designed configuration utilise char in the combustor followed by a cyclone and gas cooler. The authors of this report, while having modelled the designed configuration, strongly recommend that a coal boiler be used to either directly or indirectly heat the pyrolysis.

As a capture technology, calcium (chemical) looping is at the beginning of its technology development, at the research stage, moving into the development stage (refer to *Figure 13*).[35] As such, it is a difficult technology to cost in terms of capital and operating costs. The authors would suggest that this not be used in future base case modelling for CO_2 separation and only used as a sensitivity case.

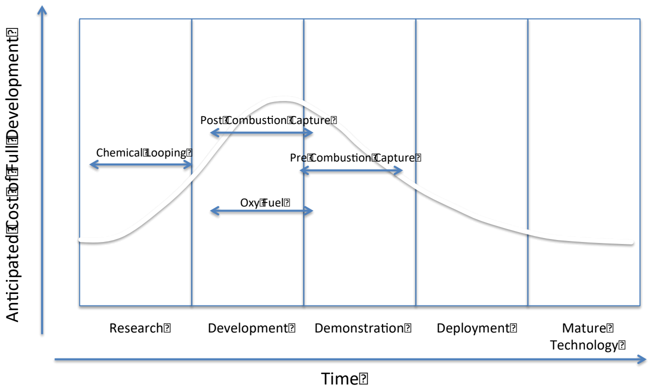

Figure 13 CO_2 capture technology development curve

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[35] Gamma Energy Technology (2018). *Power fact Book - CO_2 Technology Deployment Summary*. [online] http://www.powerfactbook.com/fact-book-chapters/carbon-dioxide-capture-technology-summary/technology-deployment-summary [Accessed Nov. 2018]

14 Summary of findings

Hydrogen offers an alternative energy option as part of a diverse future global energy market. As such, Victoria is in a unique position to become an important part of the value chain. Given its large brown coal resources, Victoria is very well-placed to become a global, cost competitive producer of hydrogen.

This study was commissioned by CSIRO as part of its project, 'Options for low cost production of CO_2-free hydrogen', co-funded by CEA and BCIA. The precursor for this work was a process simulation study, by Monash University and CSIRO, on various options for producing hydrogen from Victorian brown coal.[36] In both studies, the target hydrogen production rate was set at 770 tonnes/day, as an estimate of the quantity of hydrogen exported from Victoria to Japan in the 2030s.

The purpose of this study is to provide estimates for the likely capital cost and Levelised Cost of Product (LCoP) of hydrogen production by four different methods in 2030. The four hydrogen production options are:

1. Brown coal pyrolysis plant with tar and carbon dioxide removal by calcium based chemical looping for hydrogen production by pressure swing adsorption
2. Brown coal pyrolysis plant followed by tar removal, reforming, shift reactors and carbon dioxide removal by chemical looping for hydrogen production by pressure swing adsorption
3. Brown coal pyrolysis plant followed by tar removal, reforming, shift, tar cracking reactors and carbon dioxide removal by chemical looping for hydrogen production by pressure swing adsorption
4. Brown coal gasification plant using oxygen blown entrained for gasifier followed by shift reactor for hydrogen production by pressure swing adsorption.

The resulting cost estimates are summarised below:

Hydrogen Capital Cost and Levelised Cost of Product (2030, AUD).

Option	**2030**	
	CAPEX ($M)	LCoP ($/kg)
1: Pyrolysis only	22,000	12.30
2: Pyrolysis & reforming	11,250	5.60
3: Pyrolysis, tar cracking and reforming	10,000	5.10
4: Gasification	5,750	3.50

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[36] Kibria, M.A. and Bhattacharya, S. (2018). *Process Simulation of Hydrogen Production Options from Victorian Brown Coal.* Monash University, Clayton, Victoria

The conversion of Victorian brown coal into hydrogen offers one of the best ways to provide a cost effective, steady and reliable energy source that is not dependent on the weather. Brown coal gasification is the lowest capital approach with the lowest cost of hydrogen production. It should be noted that thus study did not include the future cost of mine rehabilitation in the cost estimation.

Part IV Supplementary Report 3

The economics of producing H_2 from electrolysis in Victoria

Jenny Hayward - CSIRO Energy

15 Introduction

As a partner to the study being undertaken looking at the techno-economic feasibility of producing hydrogen using brown coal in Victoria for export to Japan, this report aims to examine the economics of hydrogen production from dedicated renewables in Victoria for comparison purposes. This study will help to understand the competitiveness of the renewable energy derived hydrogen production technology with zero CO_2 impact.

This study considers two prospective renewable energy sources specifically available in Victoria, namely wind and solar. The energy is used in conjunction with electrolysis technology to produce the hydrogen. The study draws upon information available from various published data as well as previous work conducted for ARENA.

As part of the cost estimation, this report also provides an overview of the technology related to renewable energy from wind and solar (PV) as well as hydrogen production technology based on electrolysis. The methodology used in the cost estimation as well as basis of input data and assumptions are also included in this report.

16 Electrolysis

An electrolyser uses electricity to split water into H_2 and O_2, which can then be separated to provide a pure stream of H_2 and O_2. The advantage using an electrolyser to produce hydrogen is the source of electricity could be from the grid or from an off-grid renewable energy farm, for example. In this study, we will however consider using electricity directly generated from dedicated wind and photovoltaic farms located in Victoria.

There are several different electrolyser technologies:

- Alkaline electrolysis (AE) - Electrochemical cell that uses a potassium hydroxide electrolyte to form H_2 at the negative electrode and O_2 at the positive electrode (Hinkley, et al., 2015). This is a low cost form of electrolysis but is unsuitable for direct coupling with intermittent renewables. It also produces hydrogen at low pressures (< 30 bar).
- Proton Exchange Membrane (PEM) - Uses a proton exchange membrane and noble metal catalysts to separate H_2 and O_2 (Hinkley, et al., 2015). While it has a higher capital cost than AE, PEM is more efficient and has a higher ramp rate which makes it more suitable for direct coupling with variable renewable energy (VRE).
- Solid Oxide Electrolyser Cells (SOEC) - Uses a ceramic metal (solid oxide) and electrolyte (zirconia dioxide) to produce hydrogen. There is potential for high efficiency, however a high operating temperature (700-1000˚C) is also required (Campey, et al., 2017). This technology is still being developed, and there are no commercial systems available yet.

While there are other prospective emerging electrolyser technologies under development which may potentially offer greater performance (such as carbon assisted electrolytic process), they are however not considered here due to the low technology readiness level (TRL). In this study the focus will be on PEM electrolyser, as this technology is the best suited to coupling with VRE (Campey, et al., 2017) (Hinkley, et al., 2015) (Siemens AG, 2015).

16.1 Technical performance

PEM electrolysis has developed over the last decade to be the technology of choice for direct coupling with VRE systems, as it has the ability to cope with variations in electrical input, which AE is unable to. It produces hydrogen at a higher pressure than AE, up to 80 bar. 70 bar is the pressure needed for pipeline transport, which means that no additional and expensive mechanical compression is required with PEM electrolysis for H_2 pipeline applications. If higher pressure is required, such as for vehicle applications (~700 bar), then some mechanical compression will be needed but still less than with an AE system. PEM electrolysis produces hydrogen at the appropriate level of purity needed for fuel cells and fuel cell vehicles. This means for these applications there is no need for pressure swing absorption to further purify the hydrogen. Therefore, for the majority of expected H_2 applications, PEM electrolysis should be the lower cost and technically simpler hydrogen production method of all electrolysis technologies (Hinkley, et al., 2015).

A summary of the technical parameters for PEM electrolysis is given in Table 34.

Table 34 Summary of technical parameters for PEM electrolysis.

Technical parameter	Value and units
Electrode	Noble metals (e.g. Pt, Ir)
Current density	0.06 – 3 A/cm^2
Nominal electrolyser temperature	50-90°C
Nominal electrolyser pressure	20-80 bar
Commercial unit size	kW - 3 MWe
Energy consumption	47 – 73 kWh/kg
Minimum downturn ratio	5 – 10 %
Start up time from standby	< 10 sec
H2 purity	99.9 – 99.9999%
Stack lifetime (hours)	20,000 – 90,000

Data source: (Hinkley, et al., 2015; Guinot, Montignac, Champel, & Vannucci, 2015; Babic, Suermann, Büchi, Gubler, & Schmidt, 2017; Parra, Zhang, Bauer, & K, 2017; Bertuccioli, et al., 2014; Hydrogenics , 2017; Smolinka, Thomassen, Oyarce, & Marchal, 2016)

Research and development is ongoing into reducing the cost, increasing the stack lifetime, reducing the minimum downturn ratio and improving the efficiency (energy consumption) of PEM electrolysers. Improvements in these parameters will inevitably reduce the levelised cost of hydrogen and increase the ability of these electrolysers to work with VRE.

While the largest system is now 3 MW, it is relative easy to build larger systems by stacking smaller systems in parallel (Šarić, Dijkstra, & Haije, 2017) (Hydrogenics , 2017).

Not all of the technical parameters listed in Table 34 are required to calculate the levelised cost of hydrogen. Future performance parameters have also been estimated in several studies. The current and future technical parameters assumed in this study are shown in Table 35.

Table 35 Assumed year 2017 and 2030 technical performance parameters

Technical parameter and units	2017 value	2030 value
Energy consumption (kWh/kg H2)	58	50
Minimum downturn ratio (%)	5	4
Stack Lifetime (hours)	47,000	47,000

16.2 Economics

Several studies have reported the capital cost of the electrolysis system, stack, balance of plant (BOP) and site, installation costs etc (Guinot, Montignac, Champel, & Vannucci, 2015) (Hinkley, et al., 2015) (Parra, Zhang, Bauer, & K, 2017) (Šarić, Dijkstra, & Haije, 2017) (Babic, Suermann, Büchi, Gubler, & Schmidt, 2017) (Shaner, Atwater, Lewis, & McFarland, 2016) (Ferrero, Gamba, Lanzini, & Santarelli, 2016) (Darras, et al., 2015) (Bartela, Kotowicz, & Dubiel, 2016) (Al-Sharafi, Sahin, Ayar, & Yilbas, 2017) (Walker, van Lanen, Fowler, & Mukherjee, 2016) (Smolinka, Thomassen, Oyarce, & Marchal, 2016) (Ainscough, Peterson, & Miller, 2014).

While a range is often provided due to differences in system sizes (where smaller systems are more expensive), because this study is considering large-scale systems we have taken the lower end of the cost ranges provided in these studies. Operations and maintenance (O&M) costs have also been provided. The average costs assumed in this study are shown in Table 36 in $AUD for year 2017. The projected total capital cost at 2030 based on projections from several studies is also shown (Hinkley, et al., 2015) (Smolinka, Thomassen, Oyarce, & Marchal, 2016) (Ainscough, Peterson, & Miller, 2014).

Table 36 Assumed 2017 electrolyser cost parameters

Cost parameter and units	Value
Uninstalled capital cost electrolyser and BOP ($/kW)	2690
Installation cost (% of capital)	12
Site preparation (% of capital)	18
Contingency (% of capital)	15
Total capital cost 2017 ($/kW)	3900
Total capital cost 2030 ($/kW)	1900
O&M (% of capital)	2

17 Renewable electricity generation

For this study, it has been assumed that the electricity required to run the electrolysers will be sourced from dedicated wind or large scale PV farms under Victorian conditions. The key technical performance parameters are the capacity factors, which are described in the Methodology section. The key economic parameters are given below.

17.1 Wind farm

The capital and O&M costs and economic parameters of wind assumed for this study are shown in Table 37 .

Table 37 Wind farm economic parameters (Hayward & Graham, 2017) (IRENA, 2016)

Parameter and unit	Value
Total capital cost 2017 ($/kW)	1950
Total capital cost 2020 ($/kW)	1929
Total capital cost 2030 ($/kW)	1827
O&M (% capex)	1.5
Lifetime (years)	25
Construction period (years)	1

17.2 Large scale PV farms

The capital and O&M costs and economic parameters of large scale PV assumed for this study are shown in Table 38.

Table 38 PV farm economic parameters (Hayward & Graham, 2017) (PV Insider, 2016)

Parameter and unit	Value
Total capital cost 2017 ($/kW)	2100
Total capital cost 2020 ($/kW)	1622
Total capital cost 2030 ($/kW)	1305
O&M ($/MWh)	6.34
Lifetime (years)	25
Construction period (years)	1

18 Methodology

18.1 Electricity generation for electrolysis

It has been assumed that the electrolysers will be directly coupled to wind or large scale PV farms in several locations around Victoria and not dependent on electricity grids.

In order to meet the requirement for 770 t/d of hydrogen, several large-scale renewable energy farms and electrolysis systems would need to be constructed. A description of the methodology for working backwards from the amount of hydrogen required (770 t/d) to the wind and solar resource data is shown in Figure 14.

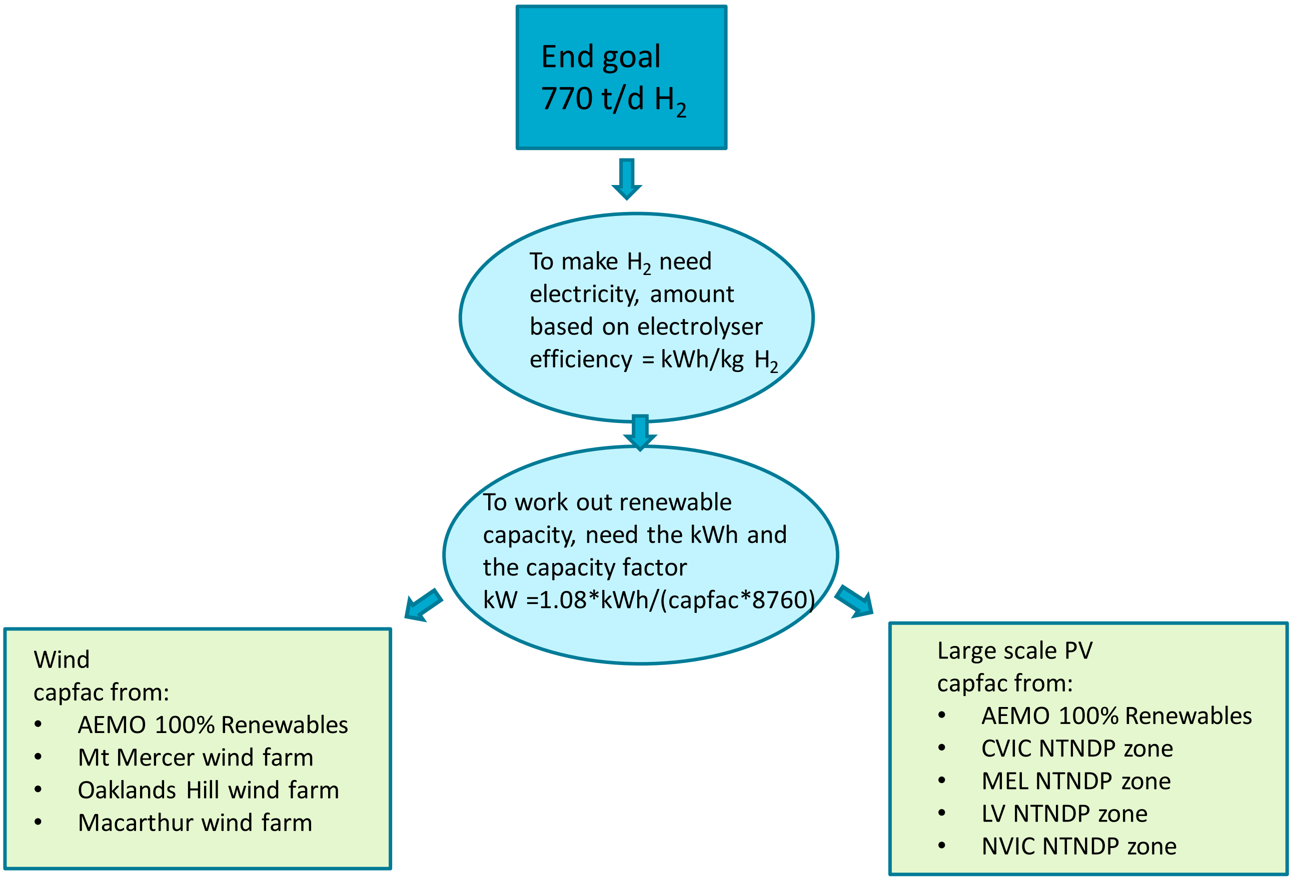


Figure 14 Simplified workflow for estimation of renewable energy requirements based on a 770 t/d of hydrogen capacity.

From the end goal, the electrolyser efficiency is used to calculate the amount of electricity required to make 770 t/d hydrogen. Then, in order to calculate the wind or large scale PV capacity required to produce this amount of electricity, the capacity factor (capfac) of wind and large scale PV is needed. In addition, since electrolysers have an auxiliary load which needs to be met before hydrogen can be generated and to take into account losses through the power electronics, it was assumed that the renewable energy farms will need to have 8% more capacity than the electrolysis system (this is shown in the diagram as a factor of 1.08).

Several sources of capacity factor were used with different values, which allowed us to calculate different wind and solar farm capacities and levelized cost of hydrogen.

Sources of wind capacity factors included:

- Victorian wind farms – Mt Mercer, Oaklands Hill and Macarthur (AEMO, 2016)
- The Australian Energy Market Operator (AEMO) 100% Renewables Study (ROAM Consulting, 2012)

Sources of large scale PV capacity factors included:

- The AEMO National Transmission Network Development Plan (NTNDP) – Country Victoria (CVIC), Melbourne (MEL), Latrobe Valley (LV) and Northern Victoria (NVIC) (AEMO, 2016)
- The AEMO 100% Renewables Study (ROAM Consulting, 2012).

18.2 The AEMO 100% Renewables Study

The AEMO 100% Renewables Study provided capacity build limits and capacity factors for wind and large scale PV farms in "polygons" around the National Energy Market (NEM) under two scenarios (ROAM Consulting, 2012). The polygons are shown in Figure 15 and it can be seen that the Victorian polygons are labelled 33 to 38b. The capacity factors are based on hourly traces of expected wind and large scale PV output over 11 years. The build limits are based on the total potential land area, taking into account land use, population density and elevation. There were differences in build limits between the two scenarios. In this study the lower, more conservative build limits were chosen (those from Scenario 2).

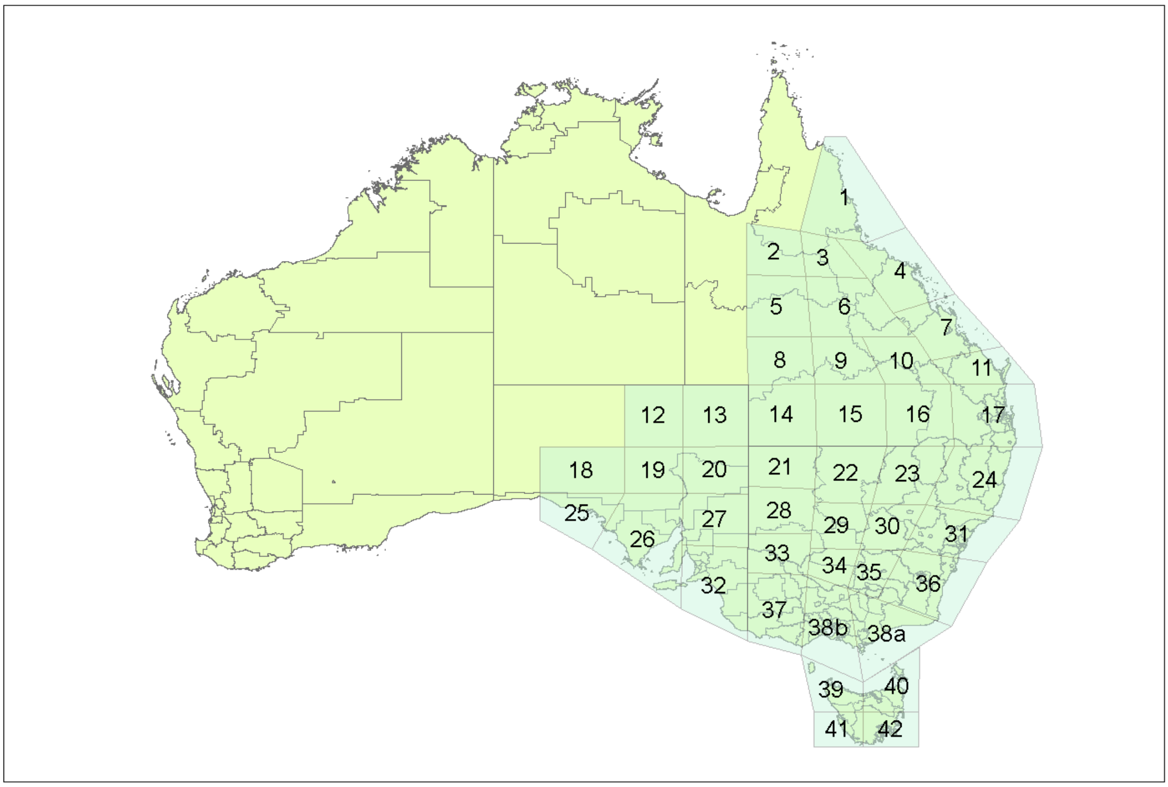


Figure 15 Polygons from AEMO 100% Renewables Study

The build limits and capacity factors for wind farms in Victorian polygons are shown in Table 39.

Table 39 Build limits and capacity factors for wind in Victorian polygons (ROAM Consulting, 2012)

Polygon number	Build limit (GW)	Capacity factor (%)
33	0	0
34	0	0
35	0.6	39.3
36	67.3	39.4
37	6.1	39.1
38a	6.8	37.6
38b	10.4	40.3

The build limits and capacity factors for large scale PV farms with single axis tracking (SAT) in Victorian polygons are shown in Table 40.

Table 40 Build limits and capacity factors for PV in Victorian polygons (ROAM Consulting, 2012)

Polygon number	Build limit (GW)	Capacity factor (%)
33	410	32.6
34	474	33.6
35	260	32.7
36	20	31.2
37	514	30.9
38a	60	30.5
38b	266	31.4

Since the polygons are located across Victoria, it was assumed that 770 t/d of hydrogen would be sourced from across all of the polygons with a capacity factor and build limit. In order to determine how much hydrogen should come from each polygon, it was assumed that at least one wind and large scale PV farm would be built in each polygon at a capacity of more than 500 MW. Then, to get to 770 t/d additional farms were assumed to be constructed in polygons with the highest capacity factors. The capacity and number of wind and large scale PV farms in each polygon is shown in the results section. The $LCOH_2$ was calculated for each individual wind and large scale PV electrolysis farm. In order to achieve a single $LCOH_2$ for Victoria from this data a weighted average was calculated according to the equation below:

$$LCOH2_{VIC} = \frac{\sum_{i=1}^{N}(LCOH2_i \times P_i)}{\sum_{i=1}^{N} P_i}$$

Where $LCOH2_{VIC}$ is the levelized cost of hydrogen for Victoria calculated using AEMO 100% Renewables data, i is the index for a farm, N is the total number of farms, $LCOH2_i$ is the levelized cost of hydrogen for an individual farm i and P_i is the hydrogen production in t/d from farm i.

18.3 Victorian wind farms

The current installed capacity and annual average capacity factors are shown in Table 41.

Table 41 Current installed capacity and capacity factors for Victorian wind farms

Wind farm	Current installed capacity (MW)	Capacity factor (%)
Oaklands Hill	63	29.0
Macarthur	420	28.8
Mt Mercer	131	29.4

These capacity factors are based on half-hourly traces measured over the 2016-2017 financial year. These values are significantly lower than that from the AEMO 100% Renewables study, by ~10%.

In order to ensure 770 t/d of hydrogen is generated from each Victorian wind farm, it was assumed that the capacity of these wind farms would increase significantly.

18.4 Victorian large scale PV farms based on the NTNDP

While there are currently no large scale PV farms in Victoria, AEMO has provided expected output data for SAT plants in the future by NTNDP zone (Table 42) (AEMO, 2016).

Table 42 Projected capacity factors for PV by NTNDP zone (AEMO, 2016)

NTNDP Zone	Capacity factor (%)
CVIC	22.6
MEL	21.7
LV	21.4
NVIC	24.4

It can be seen on comparison of Table 40 and Table 42 that the capacity factors from the AEMO 100% Renewables Study are higher than those of the more recent projections made by AEMO (AEMO, 2016). Capacity factor can have a large impact on the levelized cost, where a higher capacity factor results in a lower levelized cost (CSIRO , 2011).

19 Geographical footprint

The actual land area occupied across Victoria from both wind and PV farms and the electrolysers and associated equipment is also calculated for this study, to compare with the footprint from alternative methods of hydrogen production.

19.1 Wind farm

It is difficult to determine the physical footprint of a future wind farm given that the turbines can be placed in different arrangements which will depend on the resource and land availability etc. Turbines and their foundations have a footprint of ~100 m^2 each. Roads need to be built between turbines for maintenance. If a substation is required that can have a footprint of 1 ha (NSW Farmers, 2016). However, in this study it could be assumed there won't be a substation since the electricity is being fed to electrolysers. The land around the wind turbines can still be used by farmers. Therefore, considering only the turbine foundations, for a 500 MW wind farm with 3 MW turbines the number of turbines will be (rounding up) 167 which means a footprint of 1.67 ha or 33 m^2/MW.

There is no uniform definition of the total area of a wind power plant. A survey of wind farms in the United States (Denholm, Hand, Jackson, & Ong, 2009) suggested that there are two primary indices of land use– the infrastructure/direct impact area (or land temporarily or permanently disturbed by wind power plant development) and the total area (or overall area of the power plant as a whole). For 93 projects representing about 14 GW of proposed or installed capacity, the average *permanent direct impact* value reported was 0.3 ± 0.3 hectares/MW of capacity. For 161 projects representing about 25 GW of proposed or installed capacity, the average value for the *total project area* was about 34 ± 22 hectares/MW of capacity.

Based on this US data, a 500 MW wind farm may be expected to have a direct impact area of 150 ± 150 ha, and a total area of 17,000 ± 11,000 ha. Note that much of the total area involved remains suitable for other uses, such as grazing. For comparison, the largest capacity wind farm in Victoria, the 420 MW McArthur wind farm, occupies a total area of 5,500 ha, which is relatively compact.

19.2 Large scale PV farm

The average footprint of a SAT PV farm is estimated as 2.8 ha/MW (CO2CRC, 2015). Currently, the largest PV farm in Australia is the 102 MW Nyngan Solar Plant in NSW, which occupies an area of 250 ha (2.45 ha/MW). Thus, it is estimated that a 500 MW PV farm will require 1,400 ha. This is significantly more than the direct impact area of a wind farm of the same rated capacity.

19.3 Electrolysers

The footprint of the Siemens Silyzer electrolysis system is a shipping container (6.3 m x 3.1 m x 3.0 m) for 1.5 MW (Siemens AG, 2015). If we assume that a 500 MW system is made up of multiples of this system, then the number of shipping containers required (rounding up) will be 334 and the footprint, assuming a gap of 2m between containers, will be 1.48 ha.

20 Levelised cost of hydrogen

The formula for estimation of the levelised cost of hydrogen (LCOH$_2$) is shown in the equation below:

$$LCOH_2 = \frac{\sum_{t=1}^{25} \frac{C_t + O\&M_t}{(1+d)^t}}{\sum_{t=1}^{25} \frac{P_t}{(1+d)^t}}$$

Where C_t is the capital cost paid at time t, $O\&M_t$ is the O&M cost paid at time t, d is the discount rate and P_t is the production of hydrogen in tonnes per annum (tpa) at time t. Time is given in years.

The discount rate is assumed to be 7%, the total system lifetime 25 years and all costs are provided in AUD based in 2017 data.

21 Results

The results: $LCOH_2$ and footprint have been presented by year of data, then renewable energy type and then source of capacity factor data. As indicated the assumed input data used influence the estimated levelized cost of production.

21.1 Year 2017 and 2020 Data

The only difference between the 2017 and 2020 assumptions are the capital and O&M costs of PV and wind. This will only affect the LCOH2 and no other results.

21.1.1 Wind farm electrolysis

AEMO 100% Renewables data

In order to generate 770 t/d across Victoria, each wind farm's capacity was assumed to be 569 MW, each farm's electrolyser capacity was 526 MW and there were assumed to be 9 farms spread across Victoria. A diagram showing the number of farms in each polygon, the capacity of each farm and the quantity of hydrogen generated from each farm is shown in Figure 16.

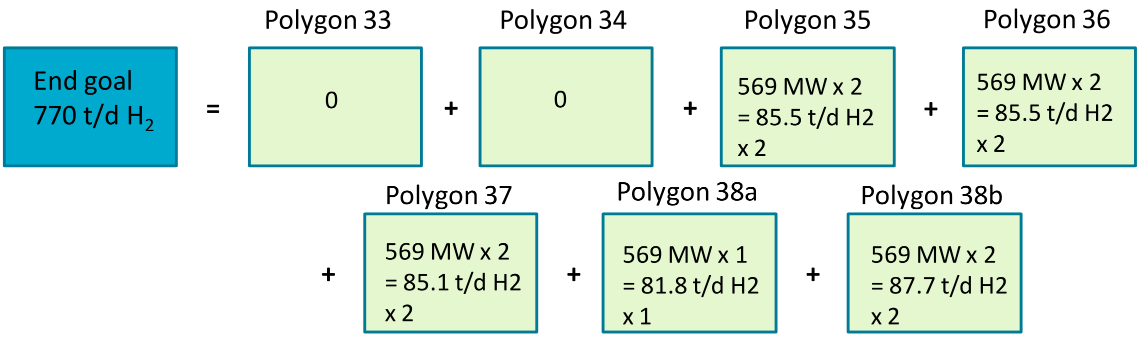


Figure 16 The size (in MW) and number of wind farms in polygons in Victoria and the amount of hydrogen generated from each farm in the years 2017 and 2020.

The results for each wind farm electrolysis system in each polygon are shown in Table 43. The weighted average $LCOH_2$ across all polygons is 10.3 $/kg in both years 2017 and 2020.

The footprint is 3.1 ha per farm (includes wind turbines, electrolysers and balance of plant) which is 27.9 ha in total across Victoria. The total Victorian wind power and electrolyser capacity is 5.1 and 4.7 GW respectively, which is the sum of all of the polygon wind (i.e. 569 MW x 9 farms) and electrolyser farms (526 MW x 9 farms) respectively.

Table 43 Results for 2017 and 2020 wind farm electrolysis using AEMO 100% Renewables data

System location	Number of farms in polygon	Hydrogen production per farm (t/d)	2017 LCOH$_2$ ($/kg)	2020 LCOH$_2$ ($/kg)
35	2	85.5	10.3	10.2
36	2	85.8	10.2	10.2
37	2	85.1	10.3	10.3
38a	1	81.8	10.7	10.7
38b	2	87.7	10.0	10.0

Mt Mercer, Oaklands Hill and Macarthur wind farm data

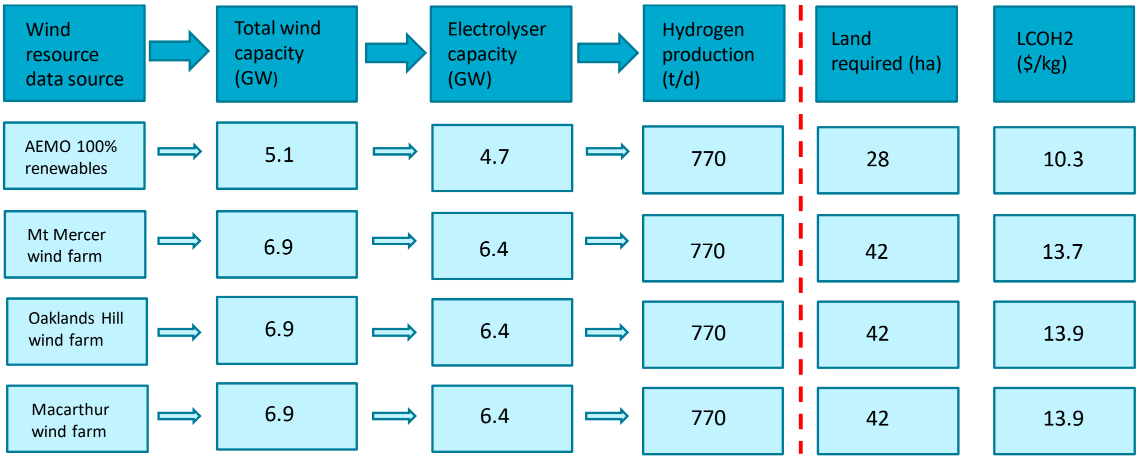
To generate 770 t/d of hydrogen using either Mt Mercer, Oaklands Hill and Macarthur capacity factors, 6.9 GW of wind and 6.4 GW of electrolyser capacity is required. The footprint of both wind and electrolysers is 42 ha. The LCOH$_2$ for the years 2017 and 2020 based on Victorian wind farm data is shown in Table 44.

Table 44 Results for 2017 and 2020 wind farm electrolysis using Victorian wind farm data

System location	2017 LCOH$_2$ ($/kg)	2020 LCOH$_2$ ($/kg)
Oaklands Hill	13.9	13.9
Macarthur	13.9	13.9
Mt Mercer	13.7	13.7

Summary

A summary of the LCOH$_2$ results from the wind farm electrolysis is shown in Figure 17.



Wind resource data source	Total wind capacity (GW)	Electrolyser capacity (GW)	Hydrogen production (t/d)	Land required (ha)	LCOH2 ($/kg)
AEMO 100% renewables	5.1	4.7	770	28	10.3
Mt Mercer wind farm	6.9	6.4	770	42	13.7
Oaklands Hill wind farm	6.9	6.4	770	42	13.9
Macarthur wind farm	6.9	6.4	770	42	13.9

Figure 17 Summary of wind electrolysis results in the years 2017 and 2020

The LCOH$_2$ is lowest when calculated using capacity factors from the AEMO 100% Renewables study. This is because the study estimated higher capacity factors than what has been observed at actual wind farms in Victoria.

21.1.2 PV farm electrolysis

AEMO 100% Renewables data

In order to generate 770 t/d of hydrogen across Victoria the capacity of each farm's electrolysis system was assumed to be 528 MW and the PV farm capacity is 570 MW (i.e. 528 Mw x 1.08) and it was assumed there were 11 farms spread across all polygons. The capacity and number of farms and the amount of hydrogen produced per farm per polygon is shown in Figure 18.

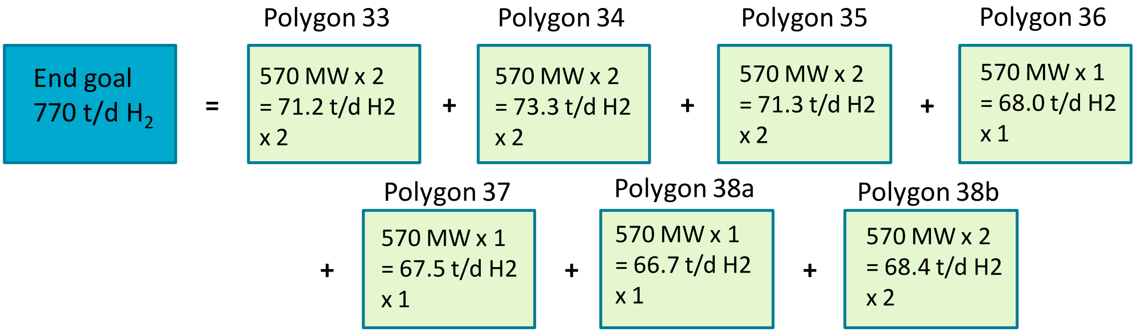


Figure 18 The size (in MW) and number of large scale PV farms in polygons in Victoria and the amount of hydrogen generated from each farm in the years 2017 and 2020

The results for each PV farm electrolysis system in each polygon are shown in Table 45.

Table 45 Results for 2017 and 2020 PV farm electrolysis calculated using AEMO 100% Renewable data

System location	Number of farms in polygon	Hydrogen production per farm (t/d)	2017 LCOH$_2$ ($/kg)	2020 LCOH$_2$ ($/kg)
33	2	71.2	12.4	11.5
34	2	73.3	12.0	11.2
35	2	71.3	12.3	11.5
36	1	68.0	12.9	12.0
37	1	67.5	12.9	12.1
38a	1	66.7	13.1	12.2
38b	2	68.4	12.8	11.9

The weighted average LCOH$_2$ across all polygons 12.5 and 11.7 $/kg in the years 2017 and 2020 respectively. This is ~2 $/kg more expensive then wind electrolysis in 2017 and this difference reduces to 1.4 $/kg by 2020. The total PV and electrolyser capacity required across Victoria calculated using the AEMO 100% Renewables data is 6.3 and 5.8 GW respectively. This is calculated as the sum of the capacities of all PV farms (570 MW x 11 farms) and electrolyser systems (528 MW x 11 farms).

The footprint for each PV electrolysis farm is 1,597.5 ha and the total for Victoria is 17,572.5 ha.

AEMO NTNDP zone data

The PV and electrolyser farm capacities and farm footprint required to achieve 770 t/d from each NTNDP zone are shown in Table 46.

Table 46 PV farm electrolysis plant capacity by NTNDP zone and total capacities required to produce 770 t/d per zone

NTNDP zone	Electrolyser capacity (GW)	PV farm capacity (GW)	Footprint (ha)
CVIC	8.2	8.9	24,978
MEL	8.6	9.3	26,010
LV	8.7	9.4	26,368
NVIC	7.6	8.3	23,123

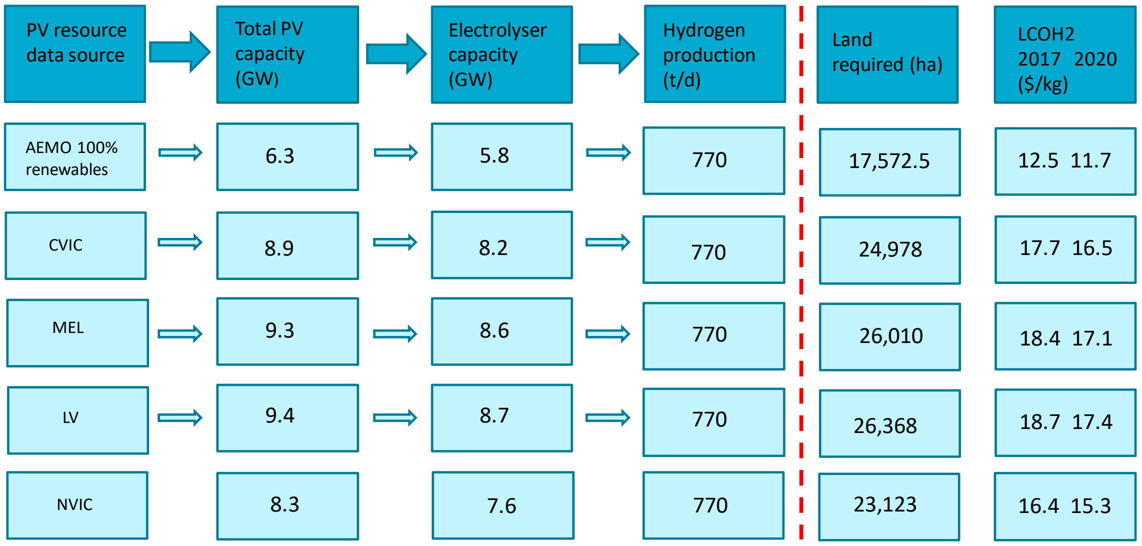
The levelized costs calculated using capacity factor data for each NTNDP zone are shown in Table 47.

Table 47 Results for 2017 and 2020 PV electrolysis using Victorian NTNDP zone data

System location	2017 $LCOH_2$ (\$/kg)	2020 $LCOH_2$ (\$/kg)
CVIC	17.7	16.5
MEL	18.4	17.1
LV	18.7	17.4
NVIC	16.4	15.3

Summary

A summary of all PV electrolysis results in the years 2017 and 2020 is shown in Figure 19.



PV resource data source	Total PV capacity (GW)	Electrolyser capacity (GW)	Hydrogen production (t/d)	Land required (ha)	LCOH2 2017 2020 ($/kg)
AEMO 100% renewables	6.3	5.8	770	17,572.5	12.5 11.7
CVIC	8.9	8.2	770	24,978	17.7 16.5
MEL	9.3	8.6	770	26,010	18.4 17.1
LV	9.4	8.7	770	26,368	18.7 17.4
NVIC	8.3	7.6	770	23,123	16.4 15.3

Figure 19 Summary of PV electrolysis results in the years 2017 and 2020

The AEMO 100% Renewables study had higher capacity factors than the NTNDP zones and thus resulted in a lower capacity, a lower footprint and lower $LCOH_2$.

21.2 Year 2030 Data

The capital and O&M costs are reduced compared to 2017 and 2020 for all systems. The electrolyser's efficiency has also improved.

21.2.1 Wind farm electrolysis

AEMO 100% Renewables data

In order to generate 770 t/d across Victoria, each wind farm's capacity was assumed to be 552 MW, each farm's electrolyser capacity was 510 MW and there were assumed to be 8 farms spread across Victoria. A diagram showing the number of farms in each polygon, the capacity of each farm and the quantity of hydrogen generated from each farm is shown in Figure 20.

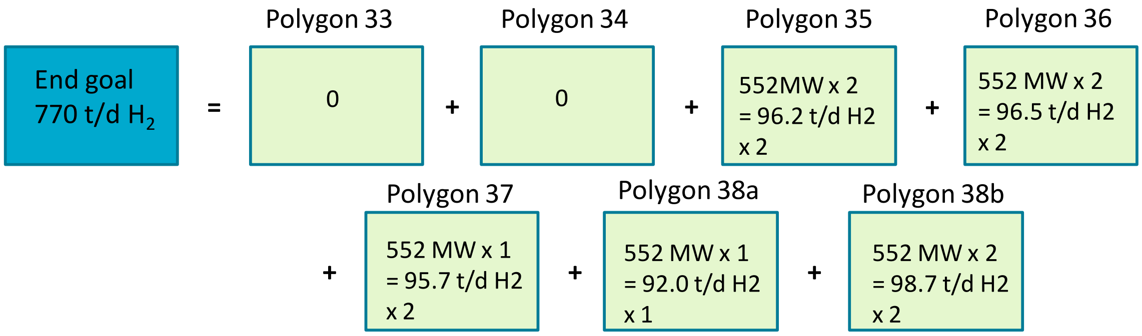


Figure 20 The size (in MW) and number of wind farms in polygons in Victoria and the amount of hydrogen generated from each farm in the year 2030

The results for each wind farm electrolysis system in each polygon are shown in Table 48.

Table 48 Results for 2030 wind farm electrolysis using AEMO 100% Renewables data

System location	Number of farms in polygon	Hydrogen production per farm (t/d)	LCOH$_2$ (\$/kg)
35	2	96.2	5.8
36	2	96.5	5.7
37	1	95.7	5.8
38a	1	92.0	6.0
38b	2	98.7	5.6

The weighted average LCOH$_2$ across all polygons is 5.8 \$/kg. The footprint is 3.3 ha per farm (wind turbines, electrolysers and balance of plant) which is 26.4 ha in total across Victoria. The total Victorian wind power and electrolyser capacity is 3.9 and 3.4 GW respectively. This is calculated as the sum of the capacities of all the wind (552 MW x 8 farms) and electrolyser (510 MW x 8 farms) capacities.

Mt Mercer, Oaklands Hill and Macarthur wind farm data

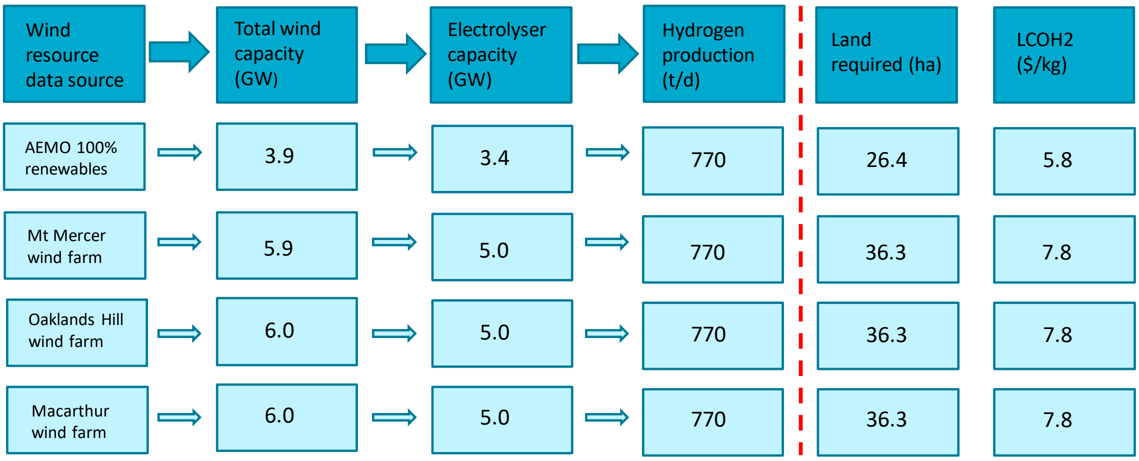
The electrolyser and wind farm capacities and footprint required to produce 770 t/d in each system location and the resultant levelized cost are shown in Table 49.

Table 49 Wind and electrolyser capacities and footprint required in 2030 to produce 770 t/d

System location	Electrolyser capacity (GW)	Wind farm capacity (GW)	Footprint (ha)	LCOH$_2$ (\$/kg)
Oaklands Hill	5.5	6.0	36.3	7.8
Macarthur	5.5	6.0	36.3	7.8
Mt Mercer	5.5	5.9	36.3	7.8

Summary

A summary of the 2030 results for wind farm electrolysis is shown in Figure 21.



Wind resource data source	Total wind capacity (GW)	Electrolyser capacity (GW)	Hydrogen production (t/d)	Land required (ha)	LCOH2 ($/kg)
AEMO 100% renewables	3.9	3.4	770	26.4	5.8
Mt Mercer wind farm	5.9	5.0	770	36.3	7.8
Oaklands Hill wind farm	6.0	5.0	770	36.3	7.8
Macarthur wind farm	6.0	5.0	770	36.3	7.8

Figure 21 Summary of wind electrolysis results in the year 2030

The levelized cost is ~5 $/kg lower than the 2017 and 2020 results. This is due to a reduction in the cost of the systems and an increase in the efficiency of electrolysers.

21.2.2 PV farm electrolysis

AEMO 100% Renewables data

In order to generate 770 t/d of hydrogen across Victoria the capacity of each farm's electrolysis system was assumed to be 499 MW and the PV farm capacity is 540 MW and it was assumed there were 10 farms spread across all polygons. The capacity and number of farms and the amount of hydrogen produced per farm per polygon is shown in Figure 22.

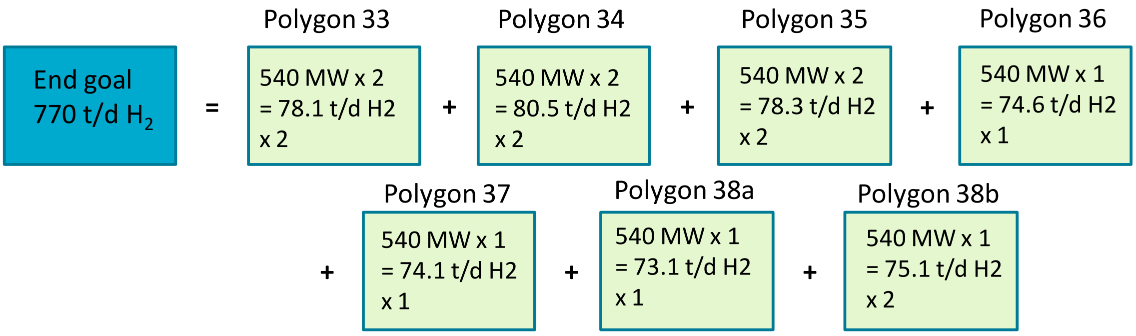


Figure 22 The size (in MW) and number of PV farms in polygons in Victoria and the amount of hydrogen generated from each farm in the year 2030

The results for each PV farm electrolysis system in each polygon are shown in Table 50. The weighted average $LCOH_2$ across all polygons is 6.0 $/kg. The total PV and electrolyser capacity across all polygons is 5.4 and 5.0 GW respectively, calculated as the sum of the capacity of all PV farms (540 MW x 10 farms) and electrolysers (499 MW x 10 farms).

The footprint for each PV electrolysis farm is 1,513.5 ha and the total for Victoria is 15,135 ha.

Table 50 Results for 2030 PV farm electrolysis calculated using AEMO 100% Renewables data

System location	Number of farms in polygon	Hydrogen production (tpd)	LCOH$_2$ (\$/kg)
33	2	78.1	5.9
34	2	80.5	5.8
35	2	78.3	5.9
36	1	74.6	6.2
37	1	74.1	6.2
38a	1	73.1	6.3
38b	1	75.1	6.1
CVIC	14	55.1	8.4
MEL	14	55.1	8.8
LV	14	55.1	8.9
NVIC	13	59.3	7.8

AEMO NTNDP zone data

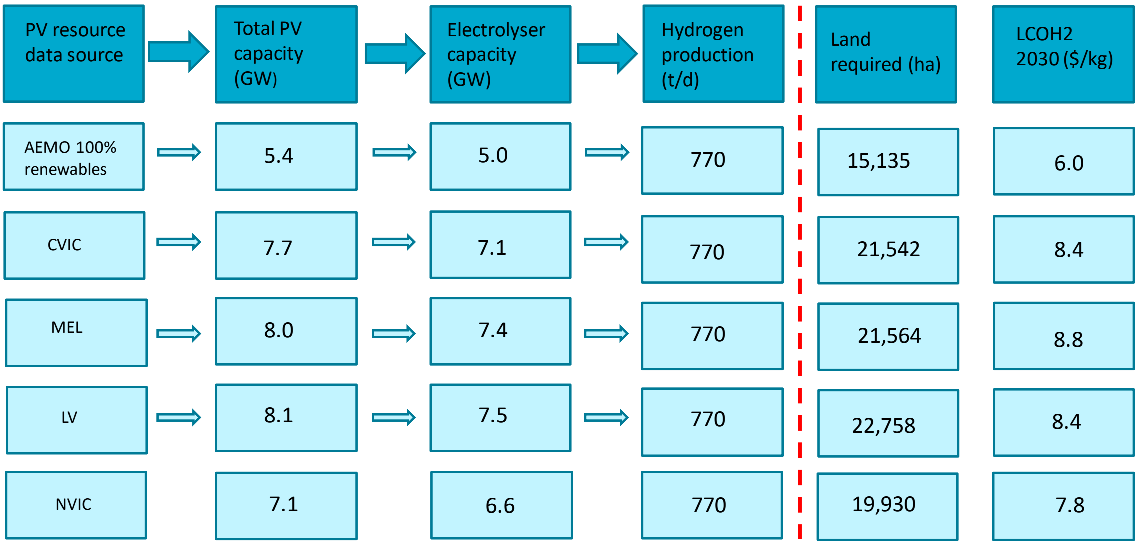
The PV and electrolyser farm capacities and farm footprint required to achieve 770 t/d and the resultant levelized cost from each NTNDP zone are shown in Table 51.

Table 51 PV farm electrolysis plant capacity by NTNDP zone

NTNDP zone	Electrolyser capacity (GW)	PV farm capacity (GW)	Footprint (ha)	LCOH$_2$ (\$/kg)
CVIC	7.1	7.7	21,542	8.4
MEL	7.4	8.0	21,564	8.8
LV	7.5	8.1	22,758	8.9
NVIC	6.6	7.1	19,930	7.8

Summary

A summary of the results for PV farm electrolysis in 2030 is shown in Figure 23.



PV resource data source	Total PV capacity (GW)	Electrolyser capacity (GW)	Hydrogen production (t/d)	Land required (ha)	LCOH2 2030 ($/kg)
AEMO 100% renewables	5.4	5.0	770	15,135	6.0
CVIC	7.7	7.1	770	21,542	8.4
MEL	8.0	7.4	770	21,564	8.8
LV	8.1	7.5	770	22,758	8.4
NVIC	7.1	6.6	770	19,930	7.8

Figure 23 Summary of PV electrolysis results in the year 2030

By comparing the AEMO 100% Renewables LCOH$_2$ in Figure 21 with Figure 23 it can be seen that PV is 0.2 $/kg more expensive then wind electrolysis in 2030. The LCOH$_2$ calculated using PV data from NVIC and calculated using Victorian wind farm data is the same (7.8 $/kg).

21.2.3 The Japanese hydrogen supply chain

The Japanese Ministry of Economy, Trade and Industry (METI) has reported that Japan will be developing commercial-scale supply chains by around the year 2030 with procurement of 300,000 tons of hydrogen annually at a cost of hydrogen around 30 yen/Nm^3.

For comparison purposes, the cost of 30 yen/Nm^3 is equivalent to about $AUD 4.10/kg H$_2$, based on a currency exchange rate of 81.50 yen per Australian dollar. The projected cost from METI is marginally lower than the cost estimated in this work.

22 Summary

As a partner to the study being undertaken looking at the techno-economic feasibility of producing hydrogen from brown coal in Victoria for export to Japan, this report aims to examine the economics of hydrogen production from dedicated renewables in Victoria. This means that a renewable energy farm is directly connected to electrolysers and not to the grid. The cost estimates exclude other cost factors such as transportation of product to the market or end-user and other infrastructure costs.

It was found that in order to meet the target of 770 t/d of hydrogen, approximately (4.4-6.9) GW of wind or (5.4-9.4) GW of photovoltaics (PV) would be required along with a similar capacity of PEM electrolysers. The different ranges are due to different years of data used. This capacity was divided into farms (of the order of around 500 or more MW capacity) and was assumed to be spread across Victoria. For a nominal capacity of 770 tonnes/day of hydrogen by electrolysis, it would need around 2 billion litres/year of purified water. The availability of suitable sites for renewable hydrogen production in term of energy and water resources will need careful consideration

The levelised cost of hydrogen production (LCOH$_2$) was calculated for the years 2017, 2020 and 2030 and improvements in the technology costs and performance were assumed to occur over time to reduce the LCOH$_2$. Different estimates of wind and solar resource data was also used to provide a range on the LCOH$_2$.

The cost of dedicated renewable energy electrolysis in Victoria is projected to decrease in cost by 2030. Wind farm electrolysis is lower cost than PV electrolysis which reflects the better wind versus solar resource in Victoria. The footprint is also significantly lower, as long as the land around the wind turbines continues to be used.

However, the gap of production cost between wind and PV becomes narrower as the capital cost of PV is assumed to be lower than wind in the years 2020 and 2030 due to technology improvement.

Using different sources for resource data, it was possible to calculate a range of LCOH$_2$, with the constraint that the total amount of hydrogen produced needs to be 770 t/d. The LCOH$_2$ results are summarised in Table 52 and the footprint data is summarised in Table 53. It is useful to note that there is no uniform definition of the total area of a wind power plant. A survey of wind farms in the United States (Denholm et al 2009) suggested that there are two primary indices of land use – the infrastructure/direct impact area (or land temporarily or permanently disturbed by wind power plant development) and the total area (or overall area of the power plant as a whole). The former approach is report in the Table 53.

Table 52 LCOH$_2$ range by year and technology

Technology	2017 LCOH$_2$ ($/kg)	2020 LCOH$_2$ ($/kg)	2030 LCOH$_2$ ($/kg)
Wind electrolysis	10.2 – 13.9	10.2 – 13.9	5.6 – 7.9
PV electrolysis	12.0 – 18.7	11.2 – 17.4	5.8 – 8.9

Table 53 Footprint by year and technology for the whole of Victoria to produce 770 tpd

Technology	2017 footprint (ha)	2020 footprint (ha)	2030 footprint (ha)
Wind electrolysis	28 - 42	28 - 42	26 - 36
PV electrolysis	17,573 – 26,368	17,573 – 26,368	15,135 – 22,758

The total size of the farms required to produce 770 t/d are shown in Table 54.

Table 54 Total renewable farm size required in whole of Victoria for 770 tpd production

Technology	2017 farm size (GW)	2020 farm size (GW)	2030 farm size (GW)
Wind electrolysis	5.1 – 6.9	5.1 – 6.9	4.4 – 6.0
PV electrolysis	6.3 – 9.4	6.3 – 9.4	5.4 – 8.1

To put these numbers into perspective, the total wind farm capacity required to produce 770 t/d is higher than Australia's total wind capacity at the end of 2016 which was 4.3 GW (Clean Energy Council, 2018). In 2016, 140 MW of wind was installed. Australia at the end of 2017 had 7 GW of PV (the majority is small scale/rooftop PV) (Australian PV Institute, 2018). This is in the mid-range of what is required to produce 770 t/d.

For comparison purpose, the projected cost from METI at 2030 (i.e. 30 yen/Nm^3 or about $AUD 4.10/kg H_2) is marginally lower than the cost estimated in this work. Like any cost estimation study, ongoing review and refinement of the cost calculation is highly recommended. For comprehensive analysis, it is required in the future to consider a range of impacts such as land use, proximity to market, transportation cost, additional investment on infrastructure etc.

23 Addendum to report

The National Hydrogen Roadmap for Australia undertaken by CSIRO has examined a base case and a best case for hydrogen production technologies, including PEM (Bruce, et al., 2018). The base case corresponds to present day technology cost and performance assumptions and the best case corresponds to future lowest cost achievable and highest performance assumptions at that price. The base case assumptions differ slightly from what has been assumed in this report. However, there are larger differences between the 2030 and best-case assumptions (or scenarios), noting that the best case may not occur until after 2030.

In light of these differences, the calculations of $LCOH_2$ have been repeated using the new assumptions for PEM electrolysers. The wind and PV cost and performance assumptions have not changed, which impacts the capacity factor of the electrolysers. This is different to what was assumed in the National Hydrogen Roadmap, where grid connected electricity was assumed to be used and green power was obtained by purchasing green energy certificates. In this study we have only looked at direct-connected wind and PV farms to electrolysers.

The key differences in assumptions are shown in Table 55.

Table 55 Comparison of assumptions used in this report and the Hydrogen Roadmap

Assumption	2017 value	Base case	2030 value	Best case
Total capital cost ($/kW)	3900	3765	1900	1005
Efficiency (kWh/kgH_2)	58	54	50	45
Stack lifetime (hours)	47,000	120,000	47,000	150,000

(Note: Base-case and Best-case are considered in the National Hydrogen Roadmap report)

Results (comparison)

The ranges in levelised cost of production of hydrogen ($LCOH_2$) are shown in Table 56. The largest differences occur between the 2030 values and the best case, where best case is ~1.5 $/kg lower in cost than the 2030 case. This is due to the lower capital cost and higher efficiency of the best case vs. the 2030 values.

Table 56 Ranges in LCOH2 calculated using assumptions from both reports

Technology	2017 $LCOH_2$ ($/kg)	Base case $LCOH_2$ ($/kg)	2030 $LCOH_2$ ($/kg)	Best case $LCOH_2$ ($/kg)
Wind electrolysis	10.2 – 13.9	9.86 – 13.7	5.6 – 7.9	4.06 – 5.67
PV electrolysis	12.0 – 18.7	11.9 – 18.4	5.8 – 8.9	3.92 – 5.99

The footprint and size of renewable farms required are 7% less under the base case compared to 2017 values and 10% less under the best case compared to the 2030 values. The comparisons are shown in Table 57 and Table 58.

Table 57 Ranges of footprint of renewable electrolysis farms calculated using assumptions from both reports

Technology	2017 footprint (ha)	Base case footprint (ha)	2030 footprint (ha)	Best case footprint (ha)
Wind electrolysis	28 - 42	26 - 39	26 - 36	23 - 32
PV electrolysis	17,573 – 26,368	16,343 – 24,522	15,135 – 22,758	13,622 – 20,482

Table 58 Ranges of capacity of renewable electrolysis farms calculated using assumptions from both reports

Technology	2017 farm size (GW)	Base case farm size (GW)	2030 farm size (GW)	Best case farm size (GW)
Wind electrolysis	5.1 – 6.9	4.7 – 6.4	4.4 – 6.0	4.0 – 5.4
PV electrolysis	6.3 – 9.4	5.9 – 8.7	5.4 – 8.1	4.9 – 7.3

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Appendix A Supplementary Report 1

Details of process simulations

1. Detailed simulation results of Option-1

The study presents a mass-energy balance for H_2 production plant using yallourn coal through pyrolysis. The production target of H_2 is 770 tons/day.

The entire plant has several sections from feed coal preparation to pyrolysis as well as downstream gas cleaning followed by pressure-swing adsorption of H_2 separation.

Several assumptions were made during simulation. The simulation begins with the mass balance from dry coal to pyrolysis product. Table 1 present the detailed mass balance of pyrolysis product and Balance between proximate and ultimate analysis of the parent coal.

Table 1: Balance of pyrolysis

SYMBOL	NAME	W%DAF	MOL%	C	H	N	S	O	ASH
DRY-CHAR	DRY-CHAR	50.6194	0.5062	46.3600	0.7277	0.4200	0.2400	0.2942	2.5800
H2O	WATER	5.5309	0.0031	0.0000	0.6189	0.0000	0.0000	4.9120	0.0000
C6H12O2	TERT-BUTYL-ACETATE	2.4253	0.0002	1.5047	0.2525	0.0000	0.0000	0.6681	0.0000
C6H6O	PHENOL	4.2426	0.0005	3.2488	0.2726	0.0000	0.0000	0.7213	0.0000
C7H8O	P-CRESOL	4.8614	0.0004	3.7797	0.3172	0.0000	0.0000	0.7646	0.0000
C10H8	NAPTHENE	1.5127	0.0001	1.4176	0.0952	0.0000	0.0000	0.0000	0.0000
H2	HYDROGEN	1.4200	0.0070	0.0000	1.4200	0.0000	0.0000	0.0000	0.0000
CH4	METHANE	3.0000	0.0019	2.2461	0.7539	0.0000	0.0000	0.0000	0.0000
C2H4	ETHYLENE	0.5830	0.0002	0.4992	0.0838	0.0000	0.0000	0.0000	0.0000
C2H6	ETAHNE	0.2271	0.0001	0.1814	0.0457	0.0000	0.0000	0.0000	0.0000
C3H6	PROPYLENE	0.0100	0.0000	0.0086	0.0014	0.0000	0.0000	0.0000	0.0000
C3H8	PROPANE	0.0100	0.0000	0.0082	0.0018	0.0000	0.0000	0.0000	0.0000
CO	CO	11.5984	0.0041	4.9735	0.0000	0.0000	0.0000	6.6250	0.0000
CO2	CO2	13.8691	0.0032	3.7851	0.0000	0.0000	0.0000	10.0840	0.0000

NH3	NH3	0.0399	0.0000	0.0000	0.0071	0.0328	0.0000	0.0000	0.0000
H2S	H2S	0.0501	0.0000	0.0000	0.0030	0.0000	0.0471	0.0000	0.0000
	TOTAL	100.0000	0.5270	68.0126	4.6008	0.4528	0.2871	24.0691	2.5800

PROXIMATE	COAL	Ultimate	coal
ASH	2.58	C	68.01
MOISTURE	5.53	H	4.60
FC	48.04	N	0.45
VOL	43.85	S	0.29
		O	24.07
		ASH	2.58

The parent coal was considered with 60% moisture content. The moisture content reduced to 12 after steam tube dryer. The moisture is further reduced to 5.53 after briquette process.

The details of the individual unit operation are as follows:

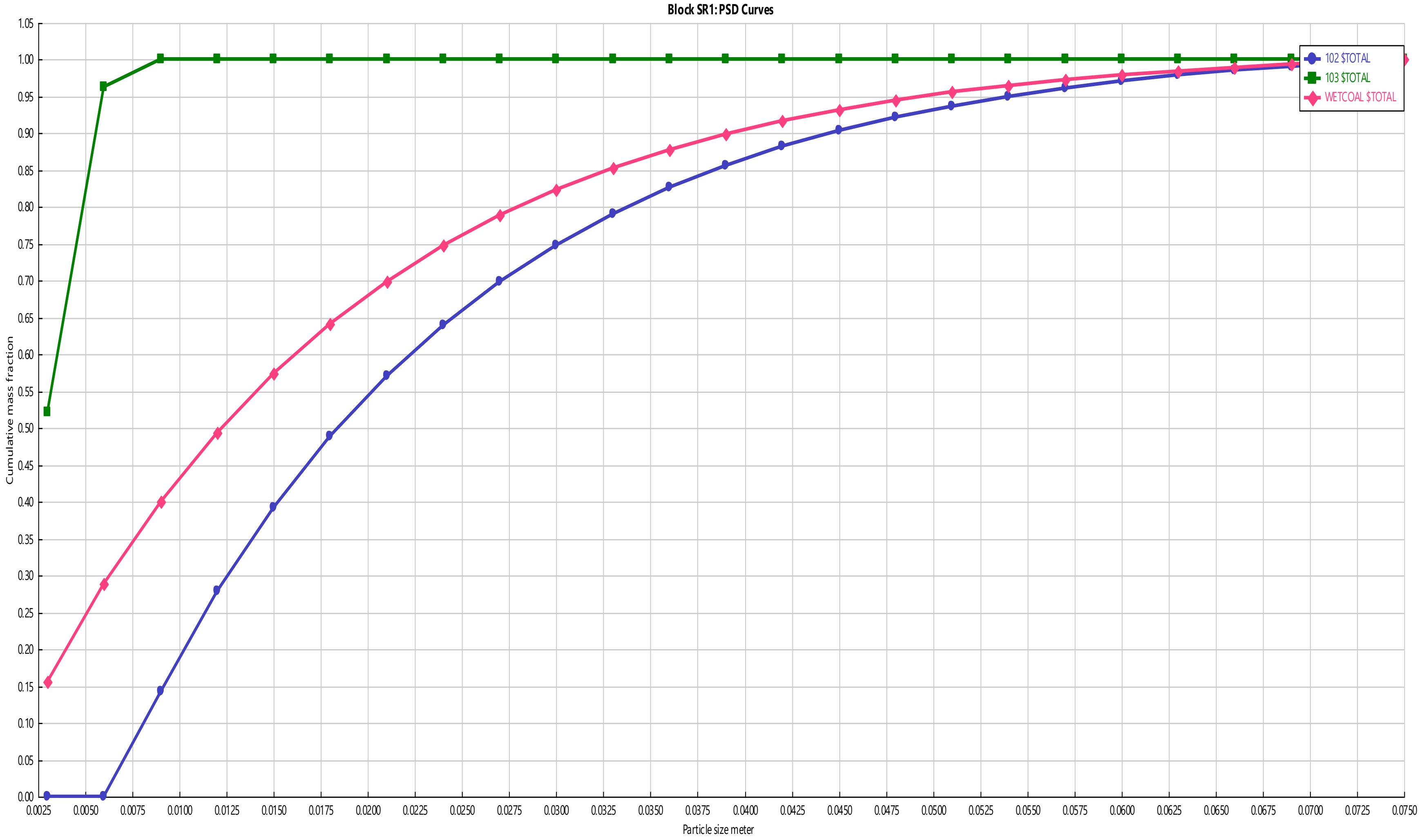
Coal storage

Coal storage is not a process operation and not modelled.

Primary screening and Hammering

The raw coal is considered with 60% moisture with a PSD as described before. This wet fuel goes to a primary hammer mill followed by a preliminary screening.

The primary screener is modelled as default SCREEN with a cut size of 0.0063 m (Coarse-split) to meet the primary Hammer mill requirement. The separation stage and efficiency is presented in **Fig 1**.


Block SR1: PSD Curves
Cumulative mass fraction
1.05
1.00
0.95
0.90
0.85
0.80
0.75
0.70
0.65
0.60
0.55
0.50
0.45
0.40
0.35
0.30
0.25
0.20
0.15
0.10
0.05
0.00
0.0025
0.0050
0.0075
0.0100
0.0125
0.0150
0.0175
0.0200
0.0225
0.0250
0.0275
0.0300
0.0325
0.0350
0.0375
0.0400
0.0425
0.0450
0.0475
0.0500
0.0525
0.0550
0.0575
0.0600
0.0625
0.0650
0.0675
0.0700
0.0725
0.0750
Particle size meter
102 $TOTAL
103 $TOTAL
WETCOAL $TOTAL

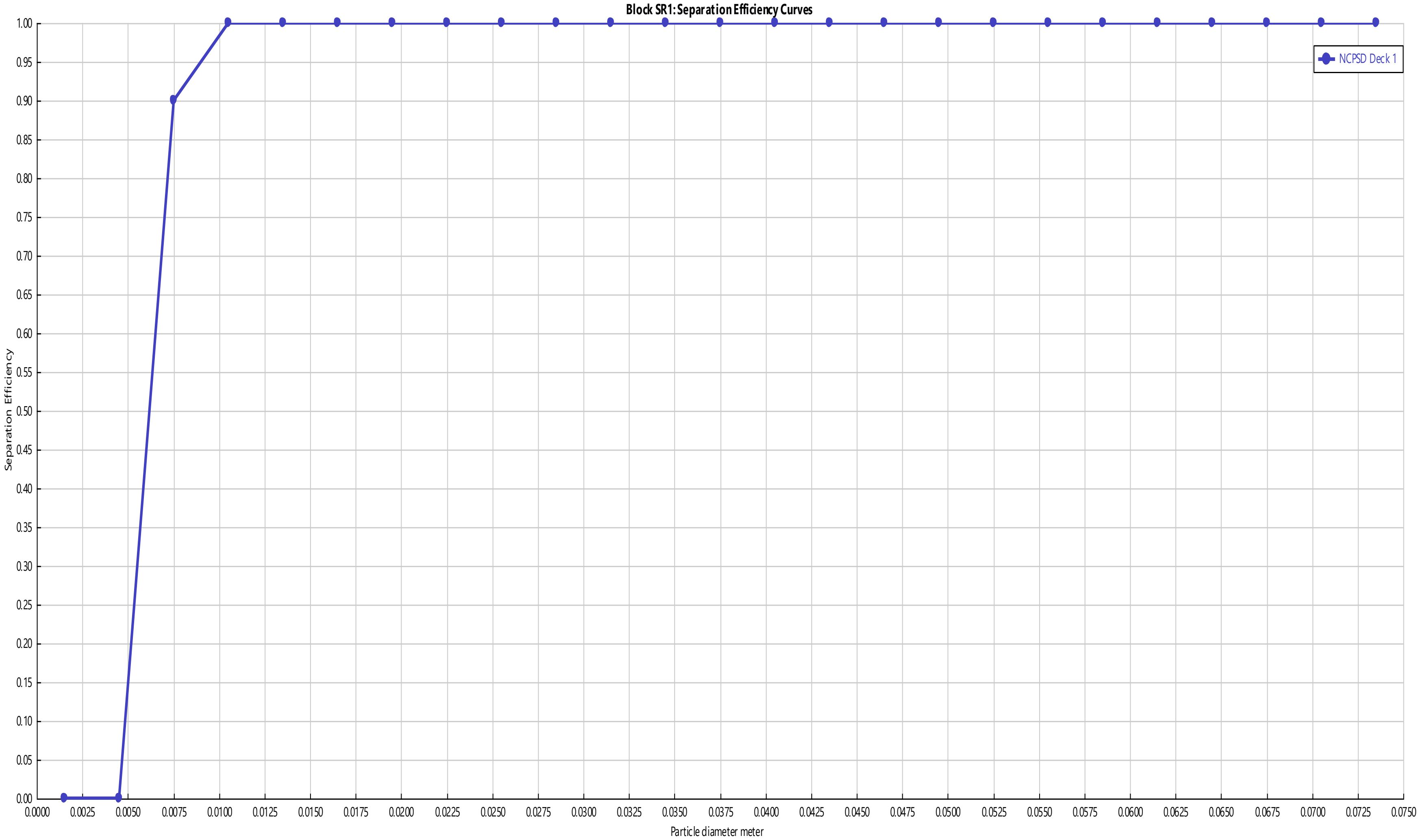
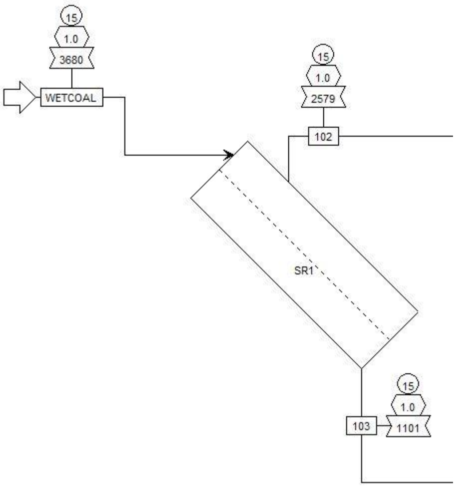
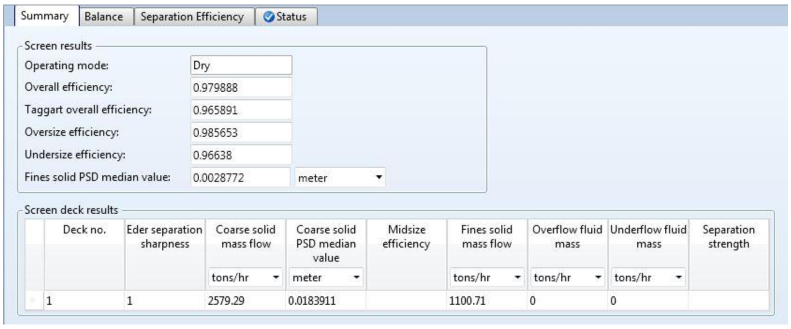


Fig 1: Primary screening separation and efficiency

The details of the primary SCREENING machine is presented in **Fig 2.**





Screen results

Operating mode: Dry
Overall efficiency: 0.979888
Taggart overall efficiency: 0.965891
Oversize efficiency: 0.985653
Undersize efficiency: 0.96638
Fines solid PSD median value: 0.0028772 meter

Screen deck results

Deck no.	Eder separation sharpness	Coarse solid mass flow tons/hr	Coarse solid PSD median value meter	Midsize efficiency	Fines solid mass flow tons/hr	Overflow fluid mass tons/hr	Underflow fluid mass tons/hr	Separation strength
1	1	2579.29	0.0183911		1100.71	0	0	

Fig 3: Details of primary screener

The hammer mill is considered as a HAMMER CRUSHER with default inbuilt model. The PSD curve shape, i.e. RRSB distribution function using dispersion parameter=2, D63 particle size adjusted to suit ~98 wt% particles <9.3 mm and secondary split. The open cut method with Bands law (HG-index 130) is used.

The PSD after primary hammering is presented in **Fig 4.**

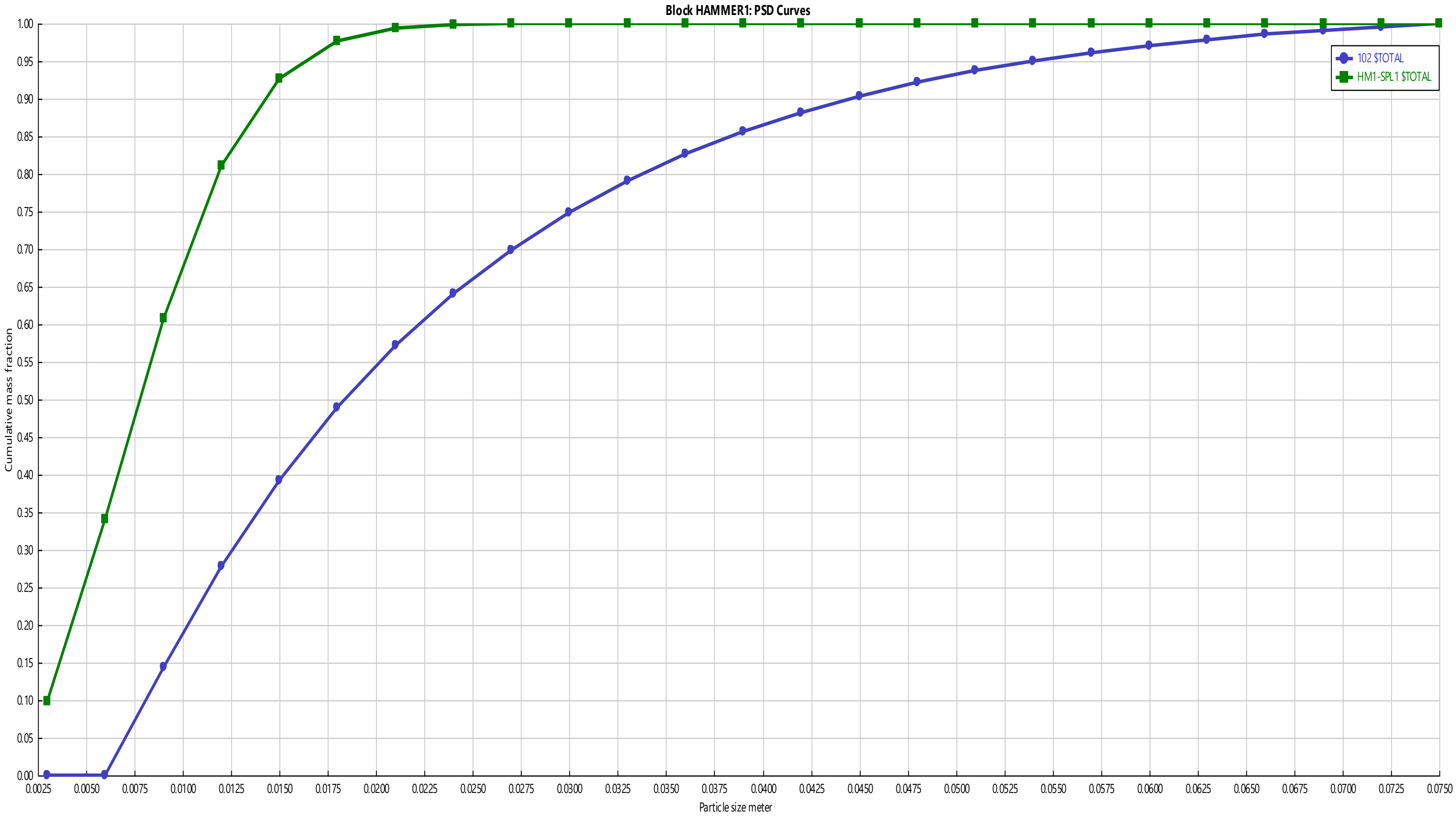
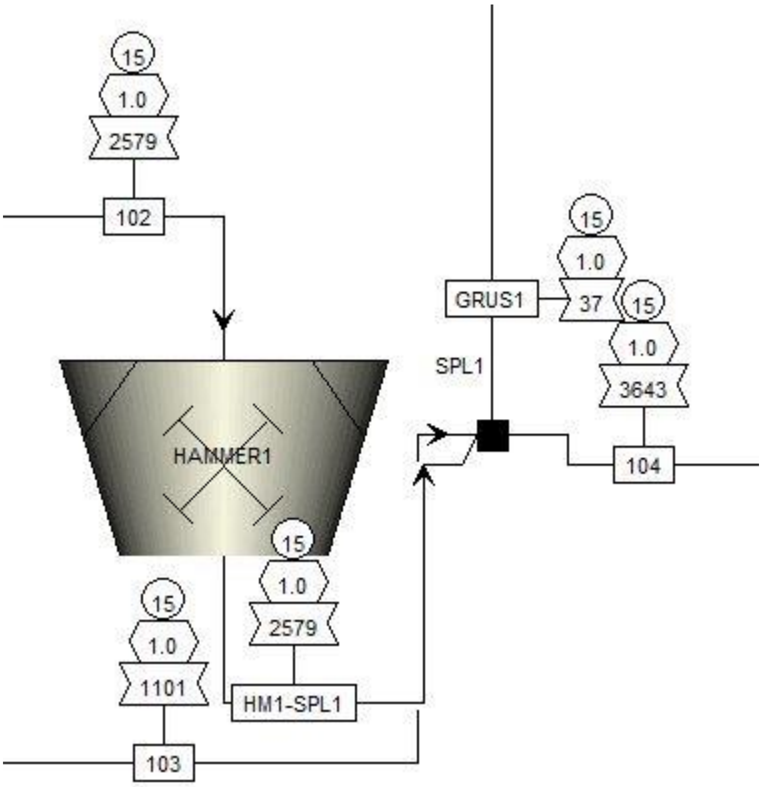
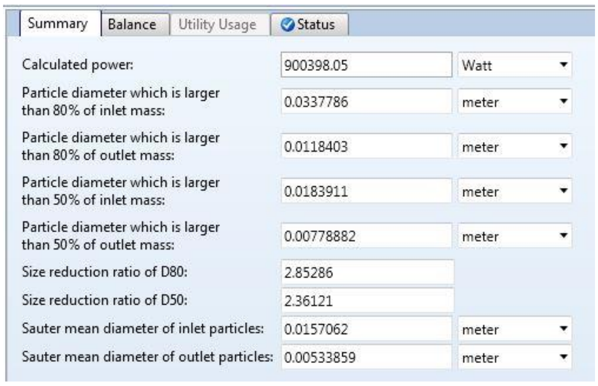


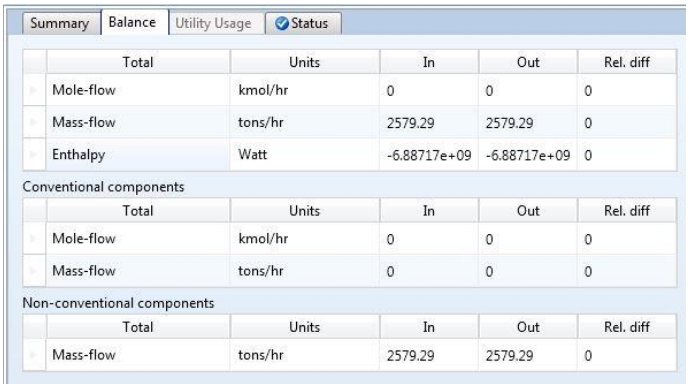
Fig 4: PSD after primary hammering

The details of the primary Hammer machine is presented in **Fig 5.**





Calculated power:	900398.05	Watt
Particle diameter which is larger than 80% of inlet mass:	0.0337786	meter
Particle diameter which is larger than 80% of outlet mass:	0.0118403	meter
Particle diameter which is larger than 50% of inlet mass:	0.0183911	meter
Particle diameter which is larger than 50% of outlet mass:	0.00778882	meter
Size reduction ratio of D80:	2.85286	
Size reduction ratio of D50:	2.36121	
Sauter mean diameter of inlet particles:	0.0157062	meter
Sauter mean diameter of outlet particles:	0.00533859	meter



Total	Units	In	Out	Rel. diff
Mole-flow	kmol/hr	0	0	0
Mass-flow	tons/hr	2579.29	2579.29	0
Enthalpy	Watt	-6.88717e+09	-6.88717e+09	0

Conventional components

Total	Units	In	Out	Rel. diff
Mole-flow	kmol/hr	0	0	0
Mass-flow	tons/hr	0	0	0

Non-conventional components

Total	Units	In	Out	Rel. diff
Mass-flow	tons/hr	2579.29	2579.29	0

Fig 5: Details of Primary Hammer

**1% of raw coal was considered as grus and was not accounted in the mainstream. An FSPLIT was used for this purpose.

Steam Tube dryer

The steam tube dryer is modelled with 3 different default inbuilt unit operation unit. An RStoic is used to convert the non-conventional moisture present in coal to convert conventional pure water at 15°C based on Eq (1):

WETCOAL--> 0.0555084 H_2O …….. (1)

The pure conventional water is passed through a heat exchanger (HX) with a specification of vapour fraction 1 at the outlet of the cold fluid. All the water that is vaporised in the HX unit is separate in a splitter (SP1). **1% of entrainment of coal was considered in the water vapour.** The overall unit of the dryer is presented in **Fig 6.**

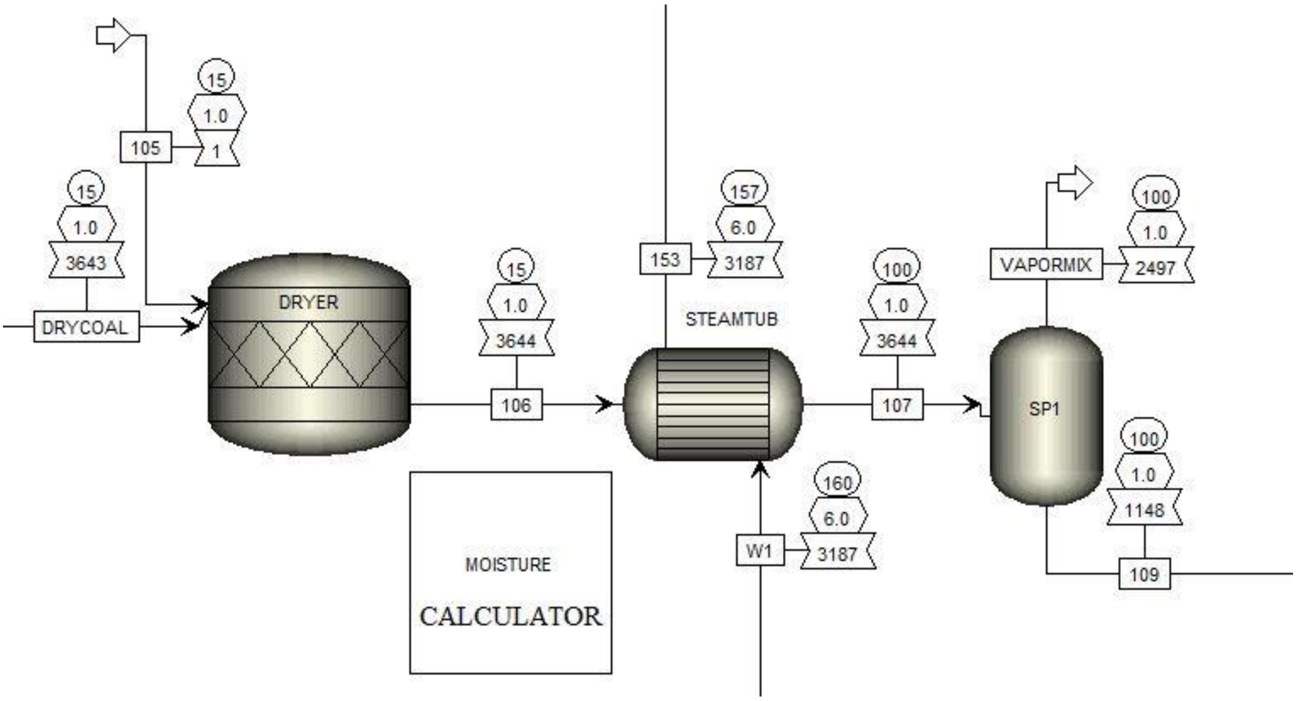
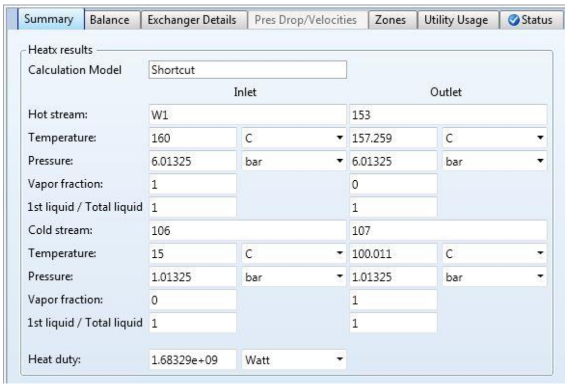


Fig 6: Steam tube dryer

A stream of air 105 is introduced in the DRYER to pull out the moisture from the tubes. 160°C and 5 barg steam were used in the hot stream of the HX. The steam requirement was adjusted such that there is 3°C temperature drop after HX. The details of the HX unit is presented in **Fig 7.**

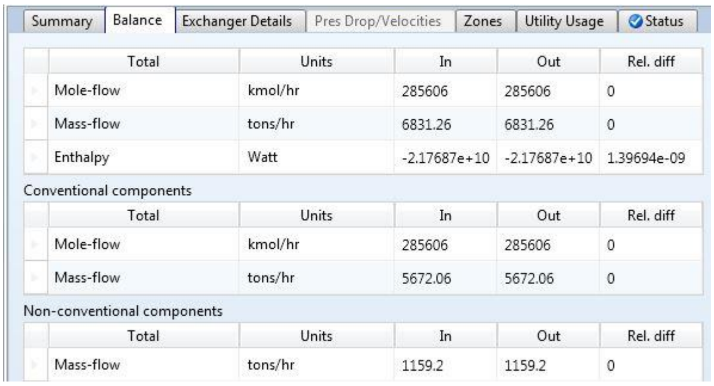


Heatx results

Calculation Model: Shortcut

	Inlet		Outlet	
Hot stream:	W1		153	
Temperature:	160	C	157.259	C
Pressure:	6.01325	bar	6.01325	bar
Vapor fraction:	1		0	
1st liquid / Total liquid	1		1	
Cold stream:	106		107	
Temperature:	15	C	100.011	C
Pressure:	1.01325	bar	1.01325	bar
Vapor fraction:	0		1	
1st liquid / Total liquid	1		1	

Heat duty: 1.68329e+09 Watt



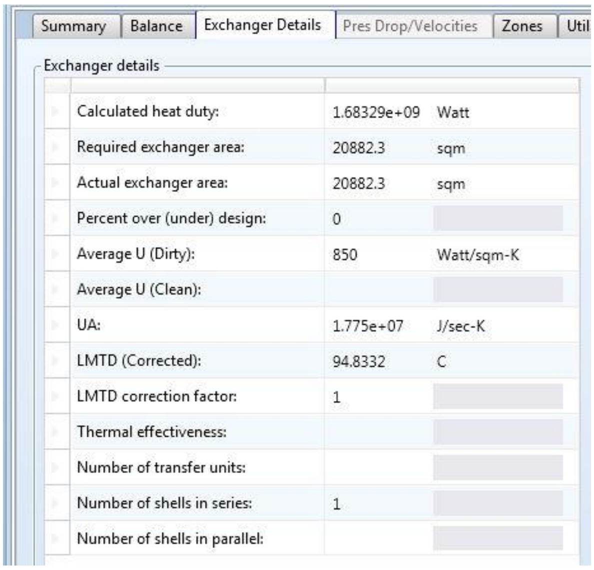
Total	Units	In	Out	Rel. diff
Mole-flow	kmol/hr	285606	285606	0
Mass-flow	tons/hr	6831.26	6831.26	0
Enthalpy	Watt	-2.17687e+10	-2.17687e+10	1.39694e-09

Conventional components

Total	Units	In	Out	Rel. diff
Mole-flow	kmol/hr	285606	285606	0
Mass-flow	tons/hr	5672.06	5672.06	0

Non-conventional components

Total	Units	In	Out	Rel. diff
Mass-flow	tons/hr	1159.2	1159.2	0



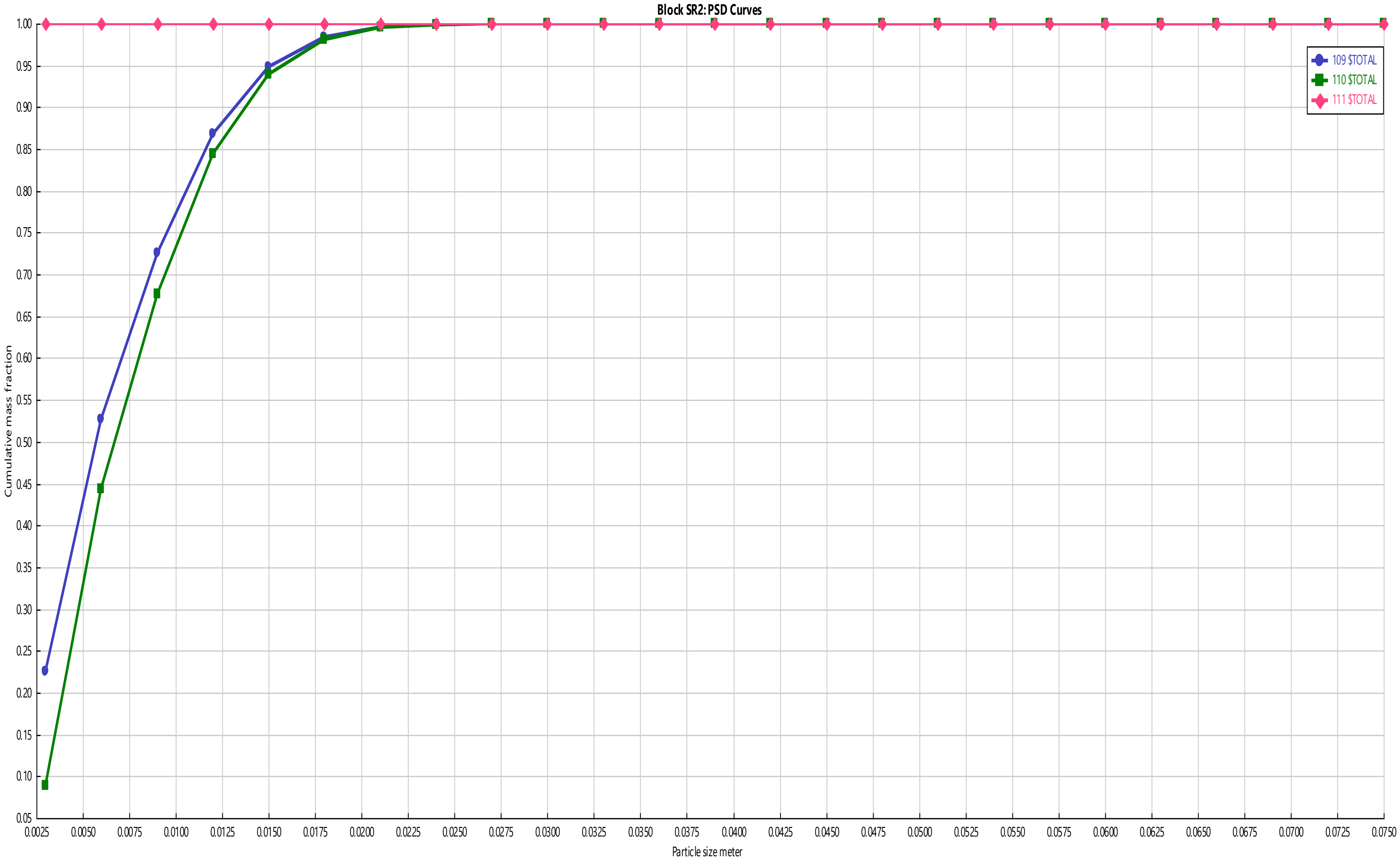
Exchanger details

Calculated heat duty:	1.68329e+09	Watt
Required exchanger area:	20882.3	sqm
Actual exchanger area:	20882.3	sqm
Percent over (under) design:	0	
Average U (Dirty):	850	Watt/sqm-K
Average U (Clean):		
UA:	1.775e+07	J/sec-K
LMTD (Corrected):	94.8332	C
LMTD correction factor:	1	
Thermal effectiveness:		
Number of transfer units:		
Number of shells in series:	1	
Number of shells in parallel:		

Fig 7: Details of Steam tube dryer unit

Secondary screening and Hammering

The dry coal goes to a secondary screening and hammering for further refinement. The secondary screener is modelled as default SCREEN with a cut size of 0.002 m (Coarse-split) to meet the secondary Hammer mill requirement. The separation stage and efficiency is presented in **Fig 8**.


Block SR2: PSD Curves
109 $TOTAL
110 $TOTAL
111 $TOTAL
Cumulative mass fraction
1.00
0.95
0.90
0.85
0.80
0.75
0.70
0.65
0.60
0.55
0.50
0.45
0.40
0.35
0.30
0.25
0.20
0.15
0.10
0.05
0.0025
0.0050
0.0075
0.0100
0.0125
0.0150
0.0175
0.0200
0.0225
0.0250
0.0275
0.0300
0.0325
0.0350
0.0375
0.0400
0.0425
0.0450
0.0475
0.0500
0.0525
0.0550
0.0575
0.0600
0.0625
0.0650
0.0675
0.0700
0.0725
0.0750
Particle size meter

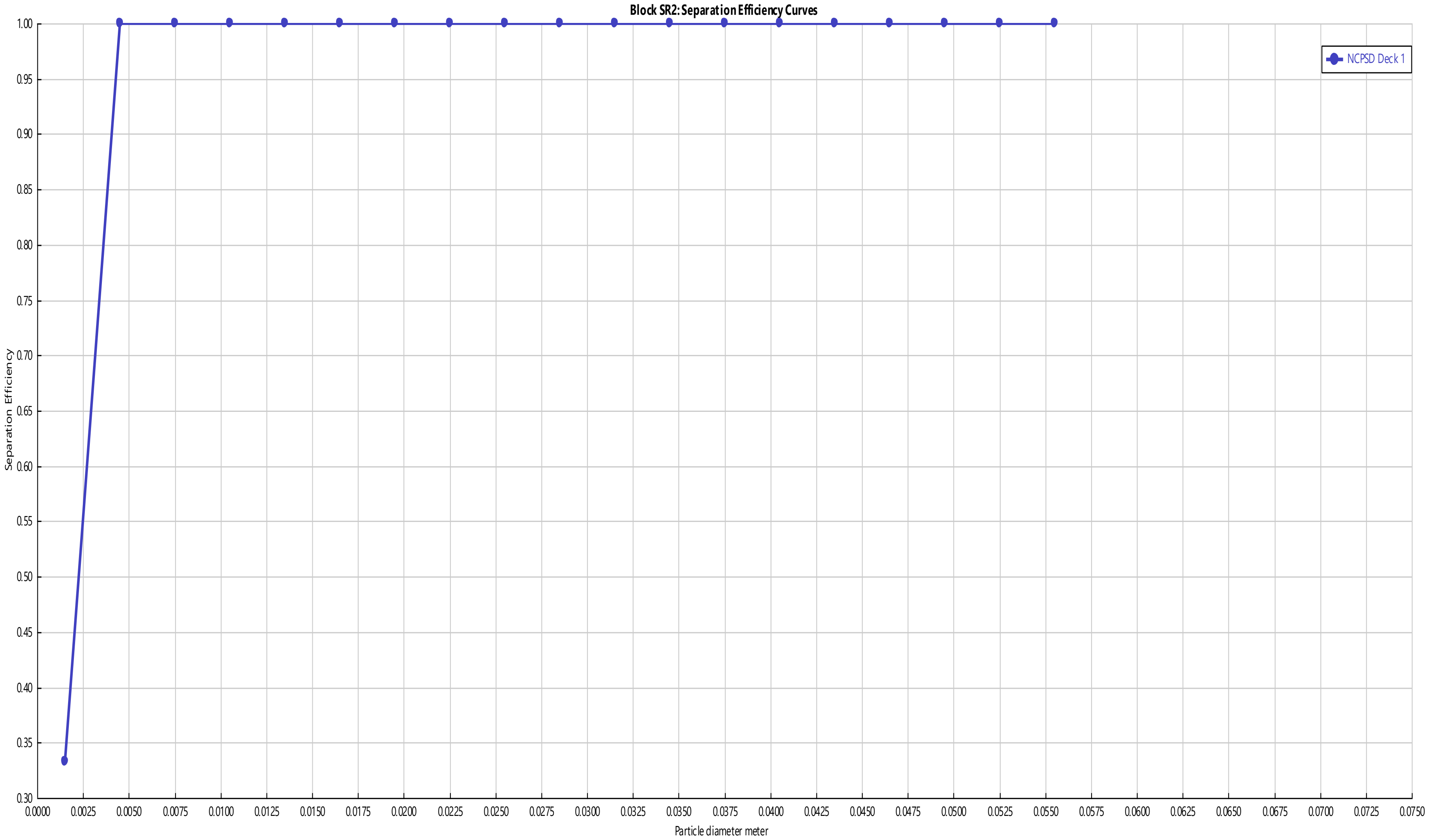
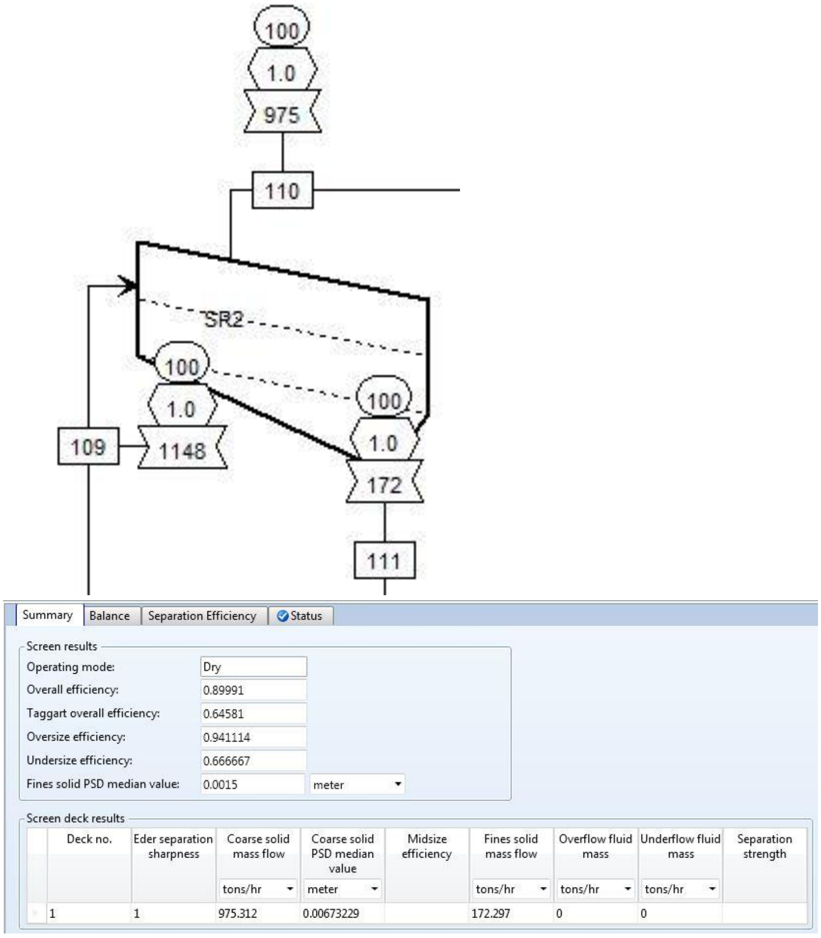


Fig 8: Secondary screening separation and efficiency

The details of the primary SCREENING machine is presented in **Fig 9.**



Screen results

Operating mode: Dry

Overall efficiency: 0.89991

Taggart overall efficiency: 0.64581

Oversize efficiency: 0.941114

Undersize efficiency: 0.666667

Fines solid PSD median value: 0.0015 meter

Screen deck results

Deck no.	Eder separation sharpness	Coarse solid mass flow tons/hr	Coarse solid PSD median value meter	Midsize efficiency	Fines solid mass flow tons/hr	Overflow fluid mass tons/hr	Underflow fluid mass tons/hr	Separation strength
1	1	975.312	0.00673229		172.297	0	0	

Fig 9: Details of secondary screener

The secondary hammer mill is considered as a HAMMER CRUSHER with default inbuilt model. The PSD curve shape i.e. RRSB distribution function using dispersion parameter=2, D63 particle size adjusted to suit ~98 wt% particles <2 mm and secondary split.

The PSD after primary hammering is presented in **Fig 10.**

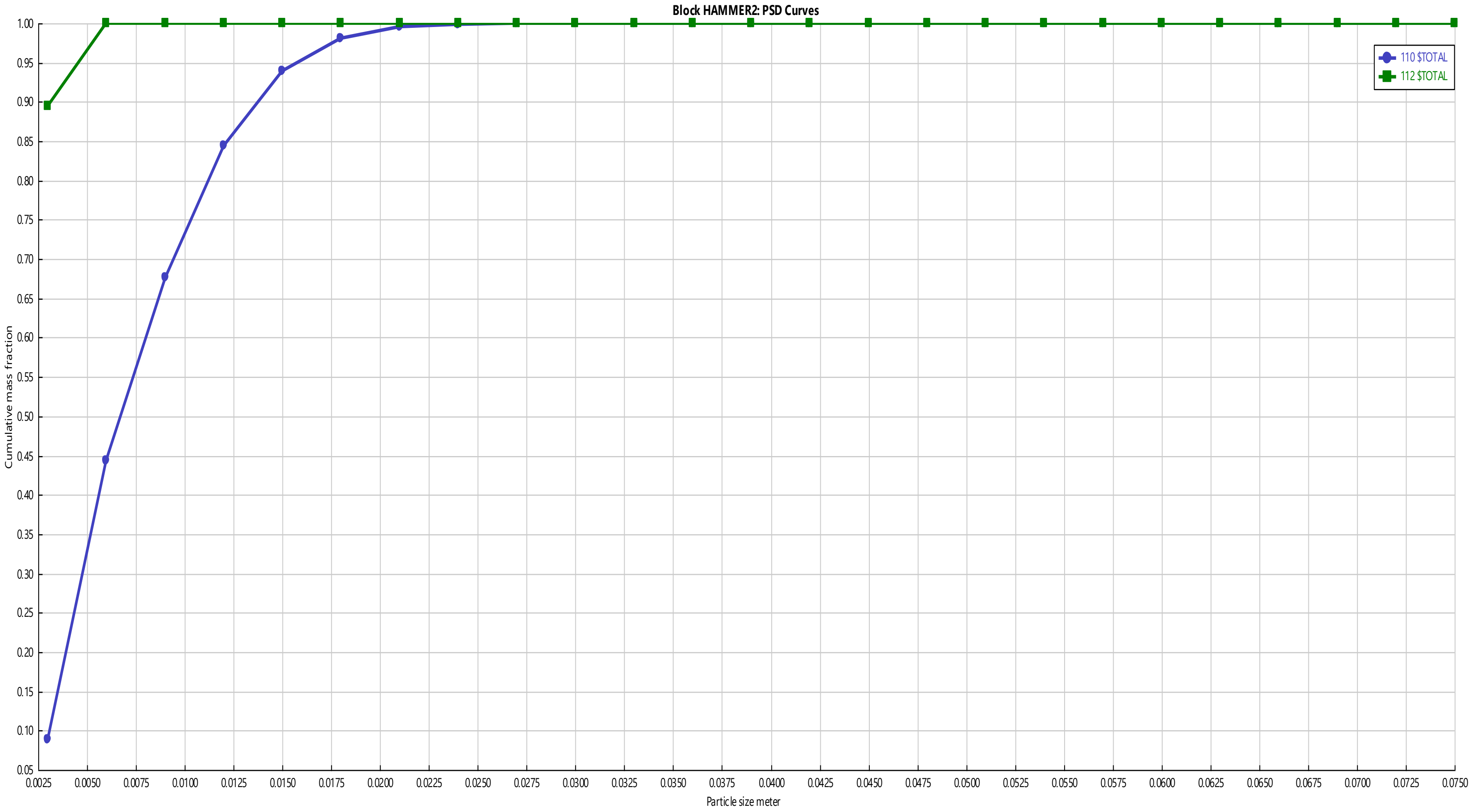
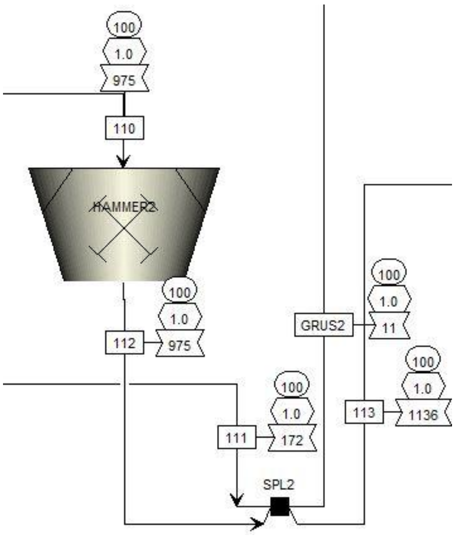
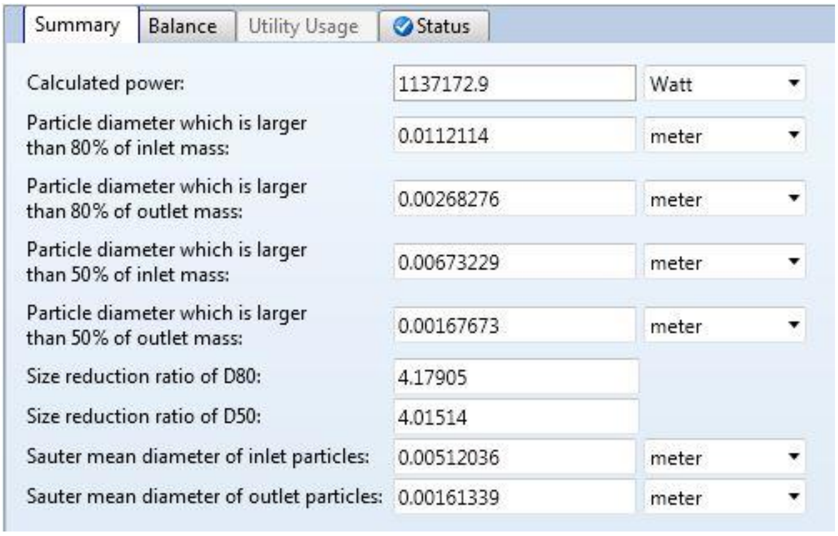


Fig 10: PSD of Secondary Hammer

The details of the secondary Hammer machine is presented in **Fig 11.**





Calculated power:	1137172.9	Watt
Particle diameter which is larger than 80% of inlet mass:	0.0112114	meter
Particle diameter which is larger than 80% of outlet mass:	0.00268276	meter
Particle diameter which is larger than 50% of inlet mass:	0.00673229	meter
Particle diameter which is larger than 50% of outlet mass:	0.00167673	meter
Size reduction ratio of D80:	4.17905	
Size reduction ratio of D50:	4.01514	
Sauter mean diameter of inlet particles:	0.00512036	meter
Sauter mean diameter of outlet particles:	0.00161339	meter

Fig 11: Details of Secondary Hammer

**1% of raw coal was considered as grus and was not accounted in the mainstream. An FSPLIT was used for this purpose.

Briquette unit

The briquette unit is modelled as default granulator model of the process simulation. The briquette is considered as 62 mm sphere diameter for an ellipsoid containing a dimension of 79x80x30 equivalent to 126 cm^3. The PSD is shown in **Fig 12** and the details in **Fig 13.**

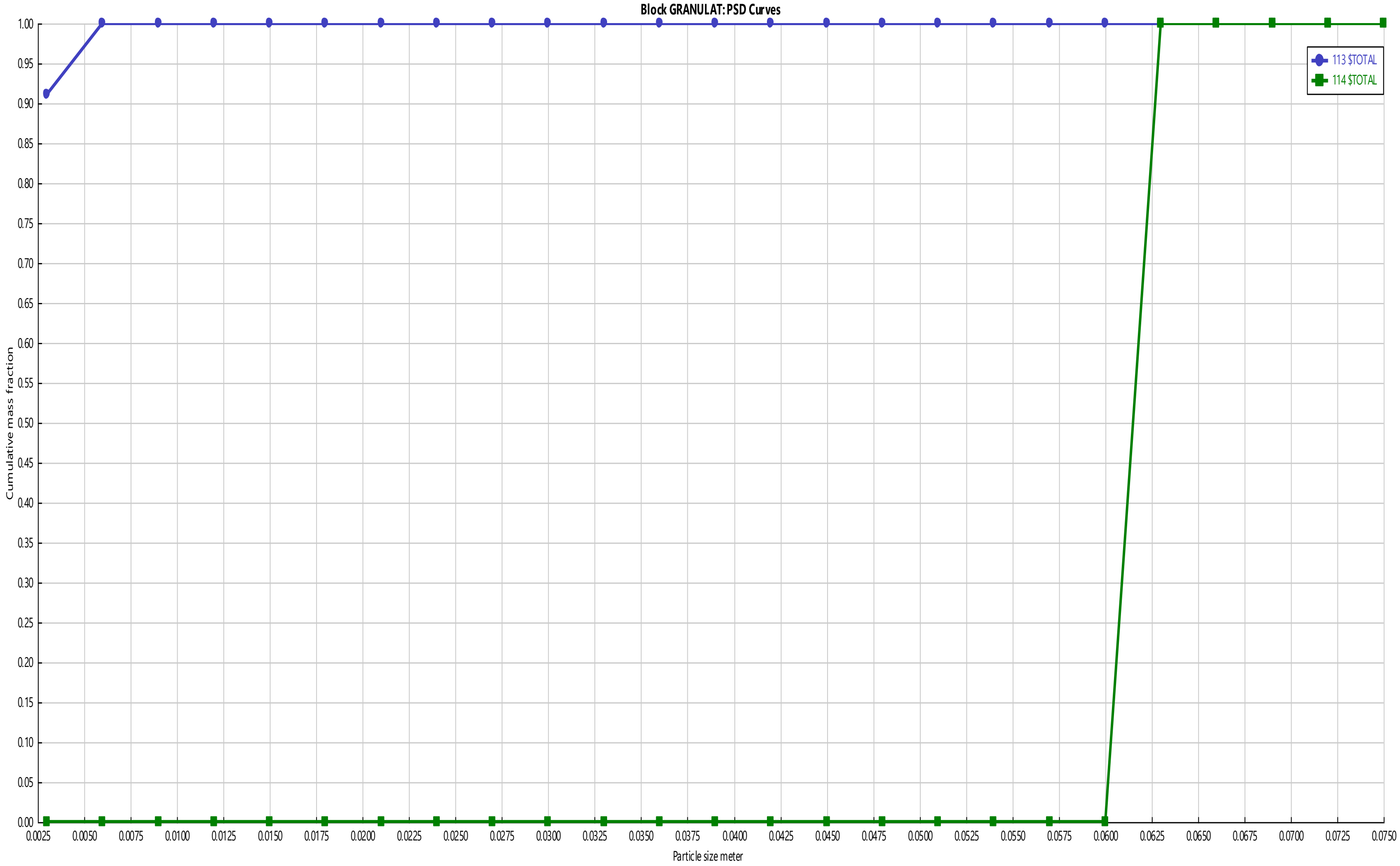
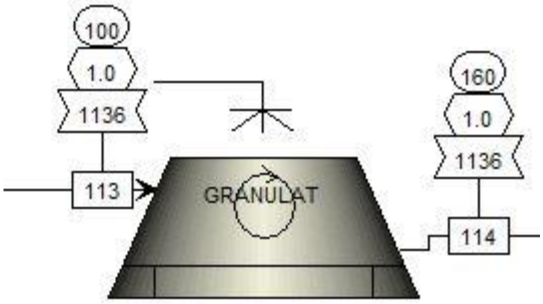
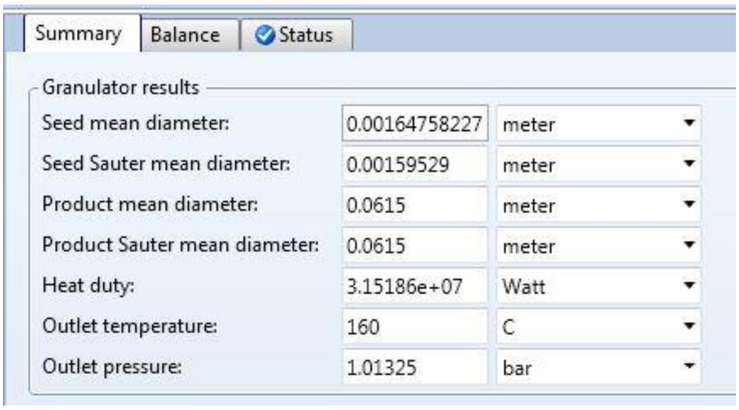


Fig 12: PSD in Granulator





Granulator results

Seed mean diameter: 0.00164758227 meter

Seed Sauter mean diameter: 0.00159529 meter

Product mean diameter: 0.0615 meter

Product Sauter mean diameter: 0.0615 meter

Heat duty: 3.15186e+07 Watt

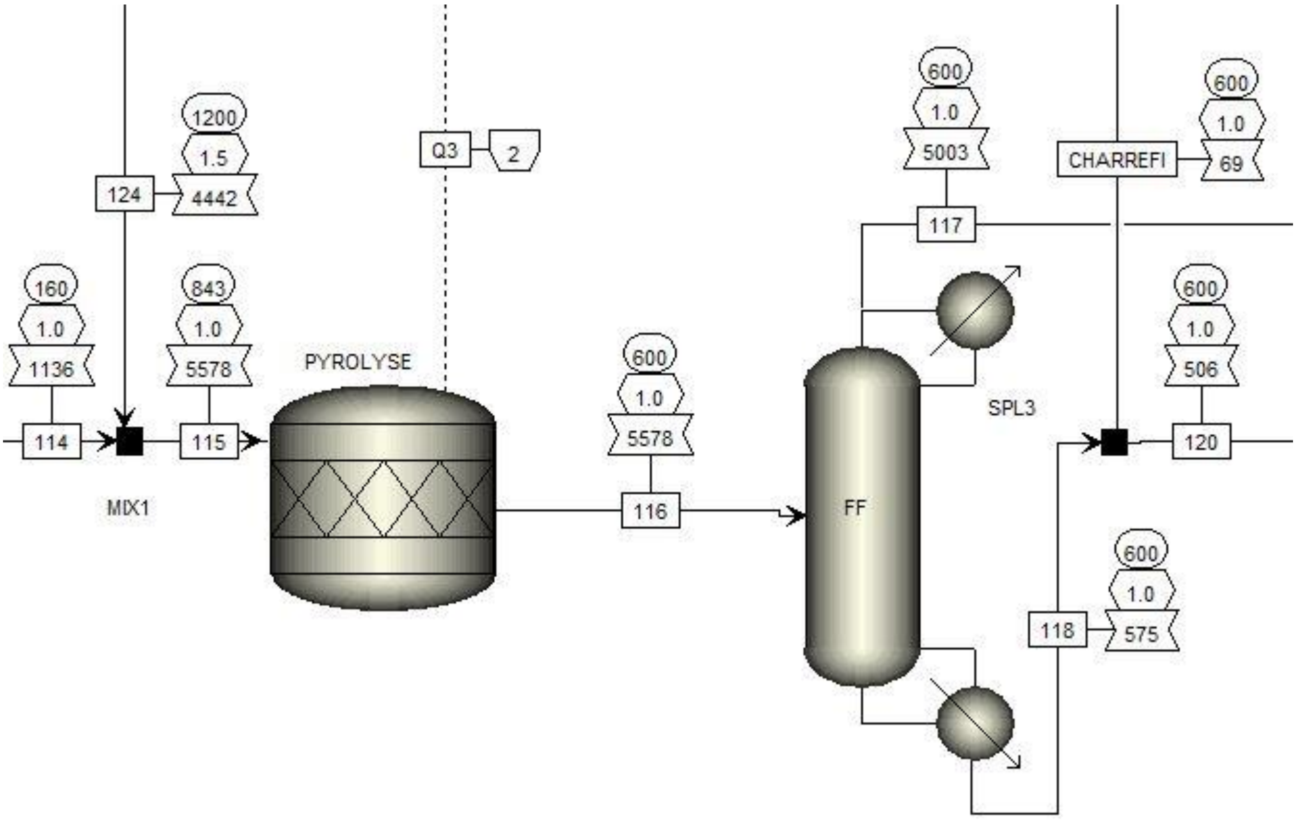
Outlet temperature: 160 C

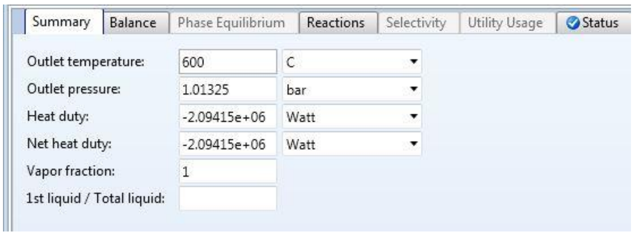
Outlet pressure: 1.01325 bar

Fig 13: Details of Granulator unit

Pyrolyser unit

The Pyrolyser unit is modelled as an RStoic reactor with a specific reactor temperature of 600°C and a component splitter. The heat requirement for the unit is satisfied by direct passing hot flue gas from the combustor. We assumed the flue gases from the combustor do not react with the pyrolysis products. The heat requirement is adjusted through the extent of char burning. It was found 88% of the char should be burnt to meet the energy requirement of the pyrolyzer. 79% of the flue gas was allowed to go inside the pyrolyzer. Excess 1% heat energy was considered in the reactor. The details of the pyrolyzer are presented in **Fig 14.**





Outlet temperature: 600 C

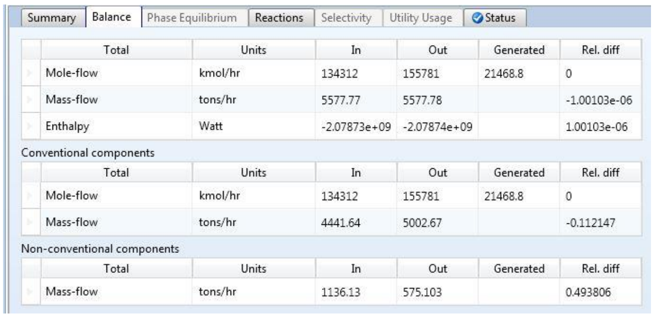
Outlet pressure: 1.01325 bar

Heat duty: -2.09415e+06 Watt

Net heat duty: -2.09415e+06 Watt

Vapor fraction: 1

1st liquid / Total liquid:



Total	Units	In	Out	Generated	Rel. diff
Mole-flow	kmol/hr	134312	155781	21468.8	0
Mass-flow	tons/hr	5577.77	5577.78		-1.00103e-06
Enthalpy	Watt	-2.07873e+09	-2.07874e+09		1.00103e-06

Conventional components

Total	Units	In	Out	Generated	Rel. diff
Mole-flow	kmol/hr	134312	155781	21468.8	0
Mass-flow	tons/hr	4441.64	5002.67		-0.112147

Non-conventional components

Total	Units	In	Out	Generated	Rel. diff
Mass-flow	tons/hr	1136.13	575.103		0.493806

Fig 14: Details of the pyrolyzer unit

Combustor and heat recovery unit

The combustor unit was modelled as an RStoic, RGibbs and a Cyclone separator presented in **Fig 15**. The RStoic is used to convert the char into its preliminary constitution. Standard combustion reactions were considered in the RGibbs reactor. The Ash was separated by Cyclone separator.

The combustor operates at 1200°C, 1.5 bar pressure. The air requirement for the char combustion was calculated with a calculator block. The air was preheated to 250°C through the hot pyrolysis gases by a heat exchanger (HX1) with a specific cold outlet temperature of 250°C. 88% of the char from the pyrolysis unit was combusted to meet the energy requirement of the pyrolyzer and other accessories. 79% of the hot flue gas was passed through pyrolyzer and the rest 21% flue gas was cooled to 300°C with a simple Heater unit to calculate the energy available to produce steam for the dryer unit.

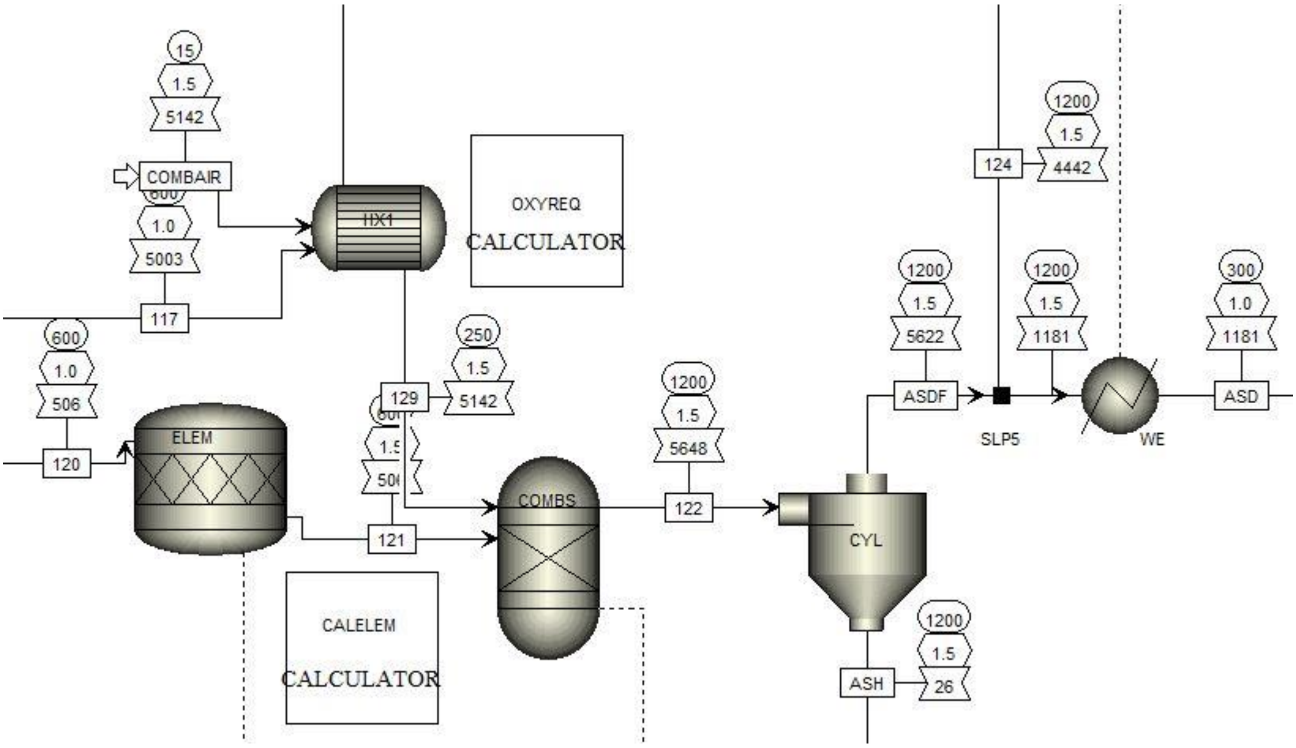
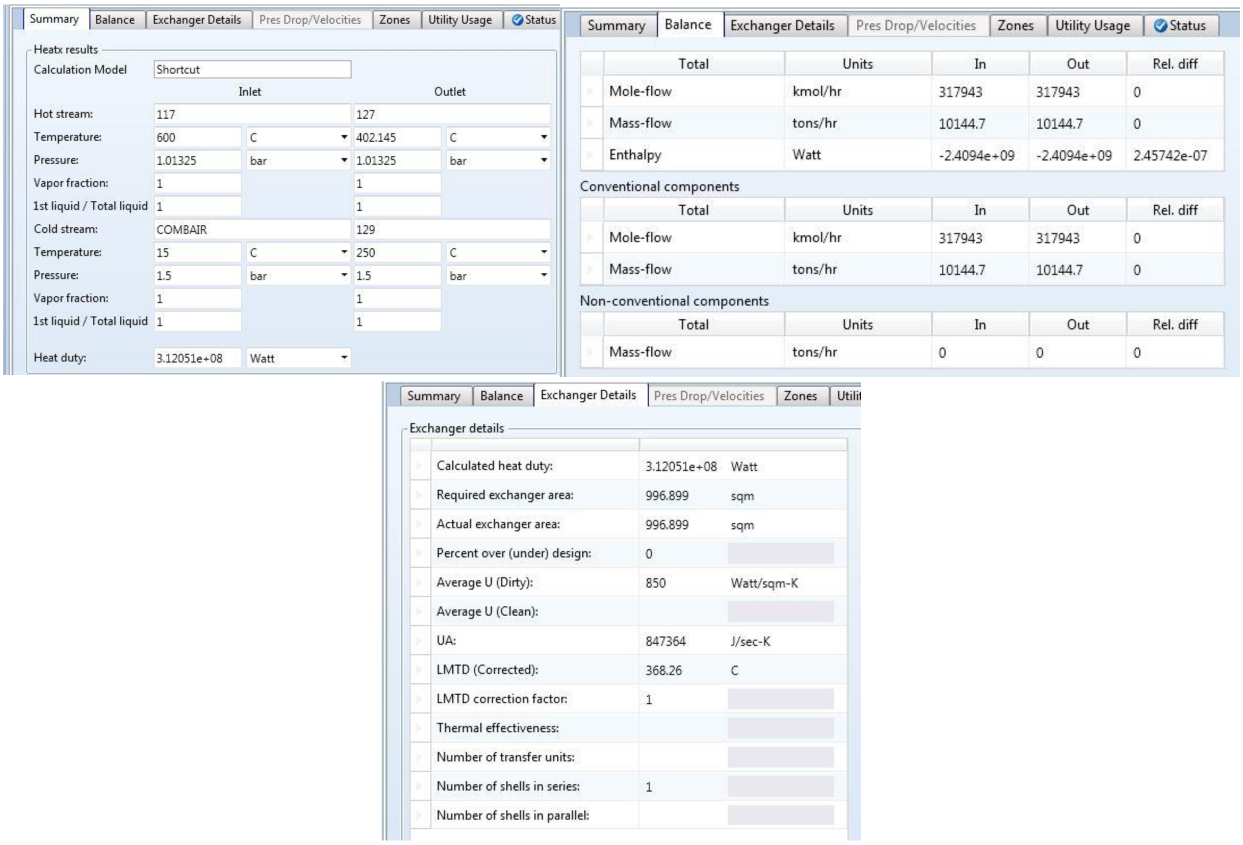


Fig 15: Combustor unit

The details of the HX1 unit is presented in **Fig 16.**



Heatx results

Calculation Model: Shortcut

	Inlet		Outlet	
Hot stream:	117		127	
Temperature:	600	C	402.145	C
Pressure:	1.01325	bar	1.01325	bar
Vapor fraction:	1		1	
1st liquid / Total liquid	1		1	
Cold stream:	COMBAIR		129	
Temperature:	15	C	250	C
Pressure:	1.5	bar	1.5	bar
Vapor fraction:	1		1	
1st liquid / Total liquid	1		1	
Heat duty:	3.12051e+08	Watt		

Total	Units	In	Out	Rel. diff
Mole-flow	kmol/hr	317943	317943	0
Mass-flow	tons/hr	10144.7	10144.7	0
Enthalpy	Watt	-2.4094e+09	-2.4094e+09	2.45742e-07

Conventional components

Total	Units	In	Out	Rel. diff
Mole-flow	kmol/hr	317943	317943	0
Mass-flow	tons/hr	10144.7	10144.7	0

Non-conventional components

Total	Units	In	Out	Rel. diff
Mass-flow	tons/hr	0	0	0

Exchanger details

Calculated heat duty:	3.12051e+08	Watt
Required exchanger area:	996.899	sqm
Actual exchanger area:	996.899	sqm
Percent over (under) design:	0	
Average U (Dirty):	850	Watt/sqm-K
Average U (Clean):		
UA:	847364	J/sec-K
LMTD (Corrected):	368.26	C
LMTD correction factor:	1	
Thermal effectiveness:		
Number of transfer units:		
Number of shells in series:	1	
Number of shells in parallel:		

Fig 16. Air preheater for char combustor

Pyrolysis Oil extraction unit

The stream of pyrolysis gas and the rest 21% flue gas were further cooled down to 25°C with a simple Heater to calculate the energy available to produce steam for the dryer unit. The liquid-vapour was separated with a FLASH unit as showed in **Fig 17**.

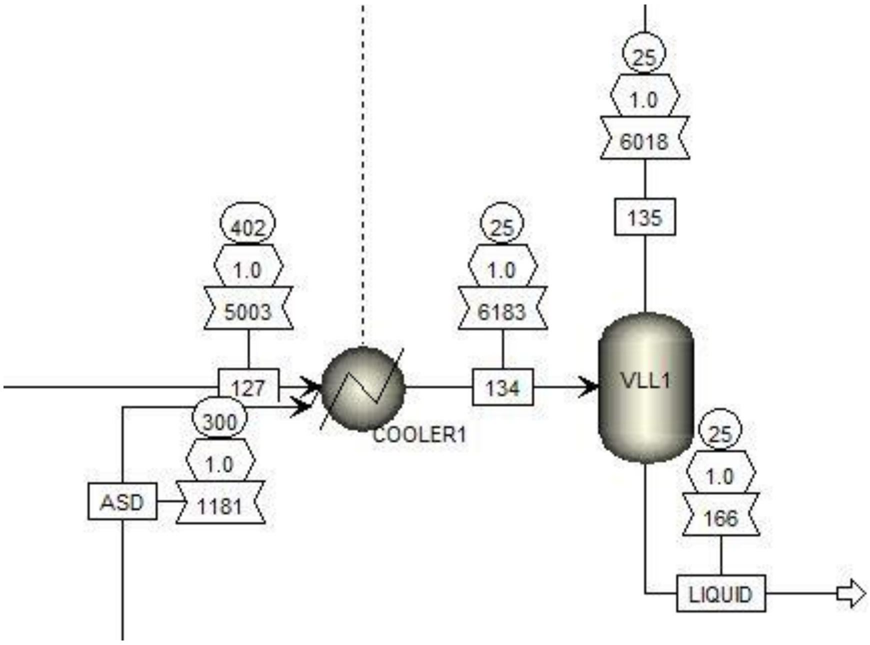


Fig 17: Pyrolysis Oil extraction unit

Gas cleaning and CO_2 separation unit

The gas cleaning and CO_2 separation unit was modelled as a $CACO_3$ looping unit presented in **Fig 18**. An extensive study was carried out, and we choose this option. $CACO_3$ looping can meet dual purposes, the pollutant removal as well as the separation of CO_2 from the mainstream. The main advantage of this unit is it does not require fresh water stream as like Amine scrubbing. There is two RGibbs reactors with two cyclone separator. In the carbonator reactor, the reaction enthalpy is exothermic and specified as 650°C. The stream of gases 135 enters the reactor and reacts with CAO.

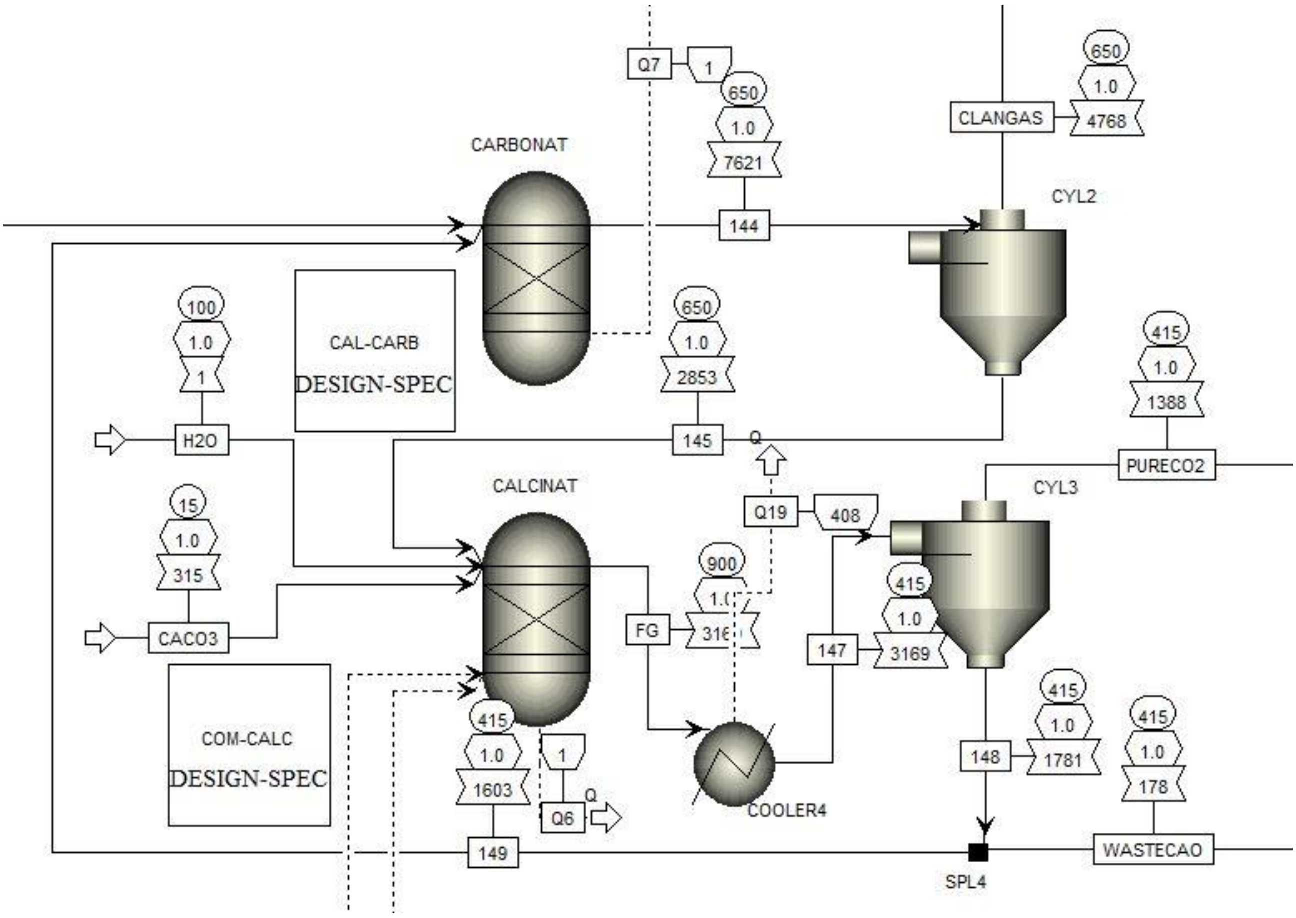


Fig 18: $CACO_3$ looping unit for gas cleaning and CO_2 separation

A design specific block was used to adjust the heat required for this block close to 1 by changing the outlet temperature of the CALCINATOR through a simple HEATER BLOCK (COOLER4).

The reaction enthalpy in the CALCINATOR is endothermic. The heat requirement of this block is satisfied by passing the excess heat from the combustor unit. A design specific block was used to adjust the feed rate of $CACO_3$ so that the total energy requirement for this block was close to 1 to a particular reactor temperature of 1000°C.

10% of the CAO was considered as waste. The entire unit was modelled in such a way so that in the recycle stream (145) there is no CAO going to CALCINATOR unit. A flow of steam (H2O) was added to the CALCINATOR so that it can drive the CO_2 and other pollutant gases from the system.

Heat recovery and PSA for H2 Separation

Energy is recovered from the clean gas and pure CO_2 stream through HEATER block as presented in **Fig 19**. The energy is then used for steam generation for the dryer. The H2 separation unit is considered as a component splitter.

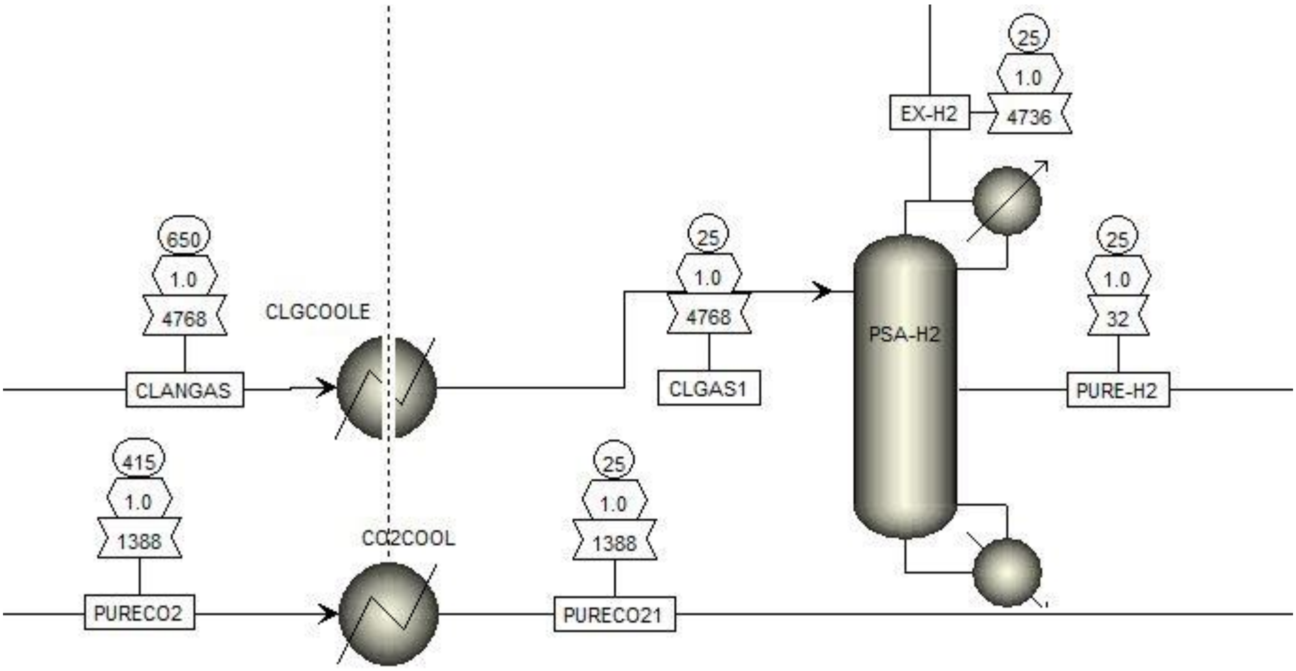


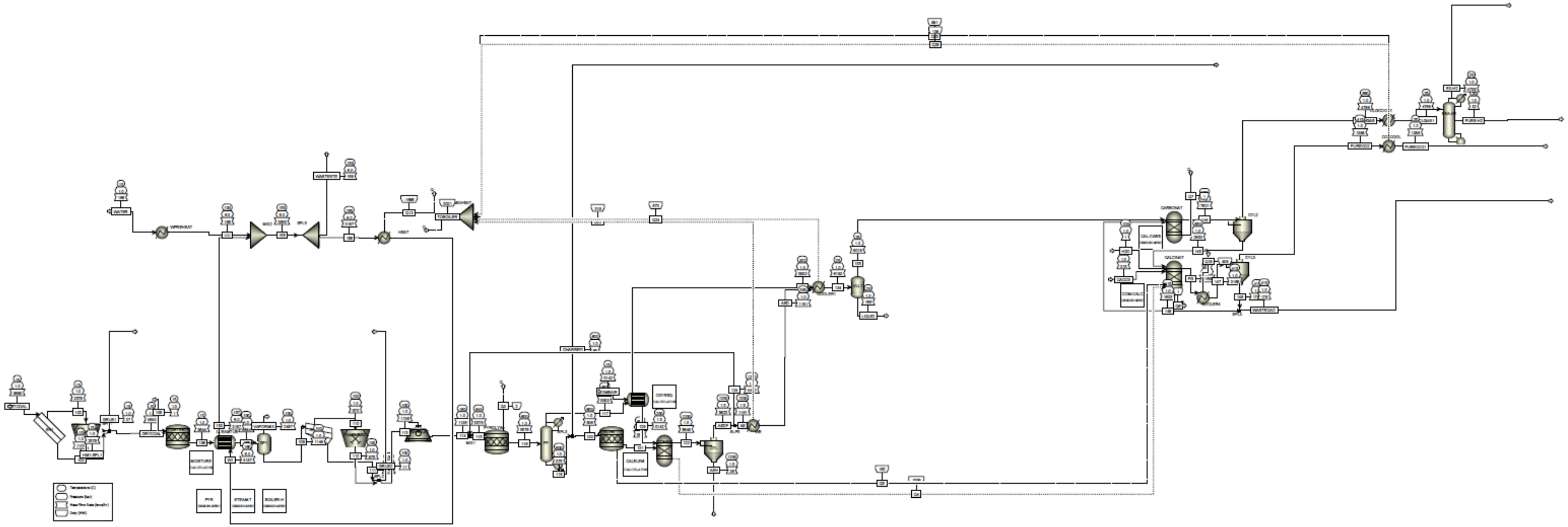
Fig 19: Heat Recovery and H2 Separation unit

Table 2 present the stream results

Table 2:

		WET COAL	STEAM-DRY	DRY-COAL	EXCESS-CHAR	LIQUID	CACO3	PURECO2	H2
Mass Flow	TONS/HR	3680	168	1136.132	69.01242	165.5851	315.3263	1388.189	32.04169
Mass Enthalpy	WATT	-9.83E+09	-673360000	-1.14E+09	37851900	-211200000	-9.59E+08	-2.76E+09	1.49E-08
Mass flow rate									
H2	TONS/HR	0	0	0	0	2.02E-05	0	0.018199	32.04169
CO	TONS/HR	0	0	0	0	0.00890894	0	187.1171	0
CO2	TONS/HR	0	0	0	0	0.3393094	0.00E+00	1200.212	0.00E+00
SO2	TONS/HR	0	0	0	0	0.0139259	0.00E+00	0.002345	0
SO3	TONS/HR	0	0	0	0	5.36E-08	0.00E+00	3.43E-11	0
NO2	TONS/HR	0	0	0	0	2.71E-14	0	0	0
NO	TONS/HR	0	0	0	0	7.00E-10	0	0	0
S	TONS/HR	0	0	0	0	0	0	0	0
H2O	TONS/HR	0	168	0	0	40.46716	0	0.835501	0.00E+00
C	TONS/HR	0	0	0	0	0	0	0	0.00E+00
C2H6	TONS/HR	0	0	0	0	0.0010356	0.00E+00	9.11E-21	0
CH4	TONS/HR	0	0	0	0	0.0015589	0.00E+00	1.02E-09	0.00E+00
H2S	TONS/HR	0	0	0	0	0.00112398	0	0.003515	0.00E+00

C3H8	TONS/HR	0	0	0	0	0.000209621	0.00E+00	2.13E-31	0.00E+00
C3H6-2	TONS/HR	0	0	0	0	0.00016325	0.00E+00	2.21E-26	0
O2	TONS/HR	0	0	0	0	3.78E-12	0.00E+00	1.48E-12	0
N2	TONS/HR	0	0	0	0	0.0693016	0	0	0.00E+00
NH3	TONS/HR	0	0	0	0	0.000762087	0	0	0.00E+00
C10H8	TONS/HR	0	0	0	0	16.79515	0.00E+00	4.70E-61	0.00E+00
C6H6O	TONS/HR	0	0	0	0	46.60445	0.00E+00	1.93E-40	0.00E+00
TERT--01	TONS/HR	0	0	0	0	6.533231	0.00E+00	1.82E-61	0.00E+00
C2H4	TONS/HR	0	0	0	0	0.00159906	0.00E+00	1.50E-16	0
P-CRE-01	TONS/HR	0	0	0	0	54.74723	0.00E+00	3.93E-51	0
CAO	TONS/HR	0	0	0	0	0	0	0	0
CACO3	TONS/HR	0	0	0	0	0	315.3263	0	0
CASO3	TONS/HR	0	0	0	0	0	0	0	0
CASO4	TONS/HR	0	0	0	0	0	0	0	0
CAS	TONS/HR	0	0	0	0	0	0	0	0
COAL	TONS/HR	3680	0	1136.132	0	0	0	0	0
CHAR	TONS/HR	0	0	0	69.01242	0	0	0	0
ASH	TONS/HR	0	0	0	0	0	0	0	0


DRYCOAL
VAPORMIX
MOISTURE
CALCULATOR
PYR
STEAM-T
BOILER-H
Temperature (C)
Pressure (bar)
Mass Flow Rate (tonne/hr)
Duty (MW)

2. Detailed simulation results of Option-2

The present study shows a mass-energy balance for H_2 production plant using yallourn coal through pyrolysis. The production target of H_2 is 770 tons/day.

The entire plant has several sections from feed coal preparation to pyrolysis as well as downstream gas cleaning followed by pressure-swing adsorption of H_2 separation.

Several assumptions were made during simulation. The simulation begins with the mass balance from dry coal to pyrolysis product. Table 3 present the detailed mass balance of pyrolysis product and Balance between proximate and ultimate analysis of the parent coal.

Table 3: Balance of pyrolysis

SYMBOL	NAME	W%DAF	MOL%	C	H	N	S	O	ASH
DRY-CHAR	DRY-CHAR	50.6194	0.5062	46.3600	0.7277	0.4200	0.2400	0.2942	2.5800
H2O	WATER	5.5309	0.0031	0.0000	0.6189	0.0000	0.0000	4.9120	0.0000
C6H12O2	2-P,4-H,M	2.4253	0.0002	1.5047	0.2525	0.0000	0.0000	0.6681	0.0000
C6H6O	PHENOL	4.2426	0.0005	3.2488	0.2726	0.0000	0.0000	0.7213	0.0000
C7H8O	P-CRESOL	4.8614	0.0004	3.7797	0.3172	0.0000	0.0000	0.7646	0.0000
C10H8	NAPTHENE	1.5127	0.0001	1.4176	0.0952	0.0000	0.0000	0.0000	0.0000
H2	HYDROGEN	1.4200	0.0070	0.0000	1.4200	0.0000	0.0000	0.0000	0.0000
CH4	METHANE	3.0000	0.0019	2.2461	0.7539	0.0000	0.0000	0.0000	0.0000
C2H4	ETHYLENE	0.5830	0.0002	0.4992	0.0838	0.0000	0.0000	0.0000	0.0000
C2H6	ETAHNE	0.2271	0.0001	0.1814	0.0457	0.0000	0.0000	0.0000	0.0000
C3H6	PROPYLENE	0.0100	0.0000	0.0086	0.0014	0.0000	0.0000	0.0000	0.0000
C3H8	PROPANE	0.0100	0.0000	0.0082	0.0018	0.0000	0.0000	0.0000	0.0000
CO	CO	11.5984	0.0041	4.9735	0.0000	0.0000	0.0000	6.6250	0.0000

CO2	CO2	13.8691	0.0032	3.7851	0.0000	0.0000	0.0000	10.0840	0.0000
NH3	NH3	0.0399	0.0000	0.0000	0.0071	0.0328	0.0000	0.0000	0.0000
H2S	H2S	0.0501	0.0000	0.0000	0.0030	0.0000	0.0471	0.0000	0.0000
	TOTAL	100.0000	0.5270	68.0126	4.6008	0.4528	0.2871	24.0691	2.5800

PROXIMATE	COAL	Ultimate	coal
ASH	2.58	C	68.01
MOISTURE	5.53	H	4.60
FC	48.04	N	0.45
VOL	43.85	S	0.29
TOTAL	100.00	O	24.07
		ASH	2.58

The parent coal was considered with 60% moisture content. The moisture content reduced to 12 after steam tube dryer. The moisture is further reduced to 5.53 after briquette process.

The details of the individual unit operation are as follows:

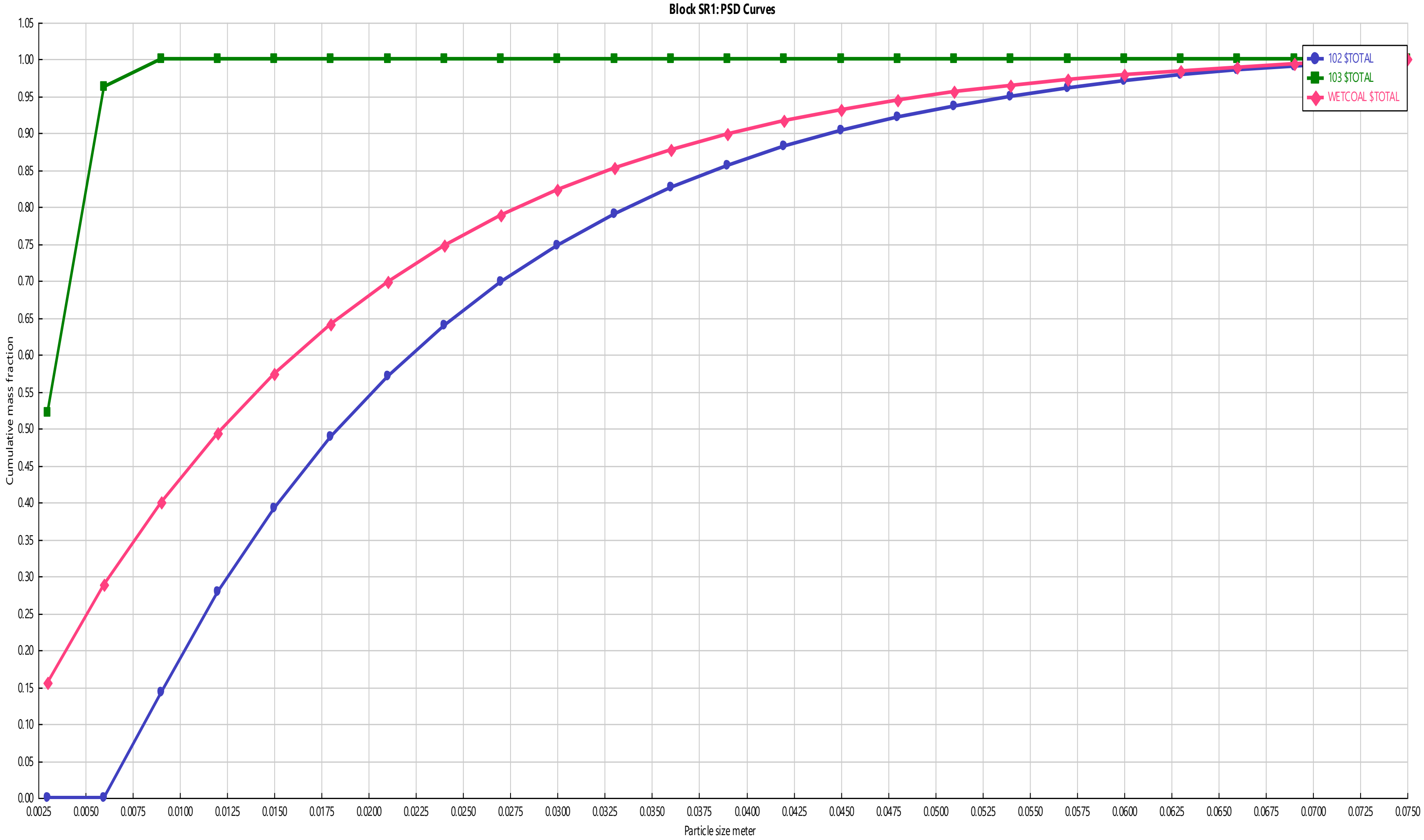
Coal storage

Coal storage is not a process operation and not modelled.

Primary screening and Hammering

The raw coal is considered with 60% moisture with a PSD as described before. This wet coal goes to a primary hammer mill followed by a preliminary screening.

The primary screener is modelled as default SCREEN with a cut size of 0.0063 m (Coarse-split) to meet the primary Hammer mill requirement. The separation stage and efficiency is presented in **Fig 20**.


Block SR1: PSD Curves
Cumulative mass fraction
1.05
1.00
0.95
0.90
0.85
0.80
0.75
0.70
0.65
0.60
0.55
0.50
0.45
0.40
0.35
0.30
0.25
0.20
0.15
0.10
0.05
0.00
0.0025
0.0050
0.0075
0.0100
0.0125
0.0150
0.0175
0.0200
0.0225
0.0250
0.0275
0.0300
0.0325
0.0350
0.0375
0.0400
0.0425
0.0450
0.0475
0.0500
0.0525
0.0550
0.0575
0.0600
0.0625
0.0650
0.0675
0.0700
0.0725
0.0750
Particle size meter
102 $TOTAL
103 $TOTAL
WETCOAL $TOTAL

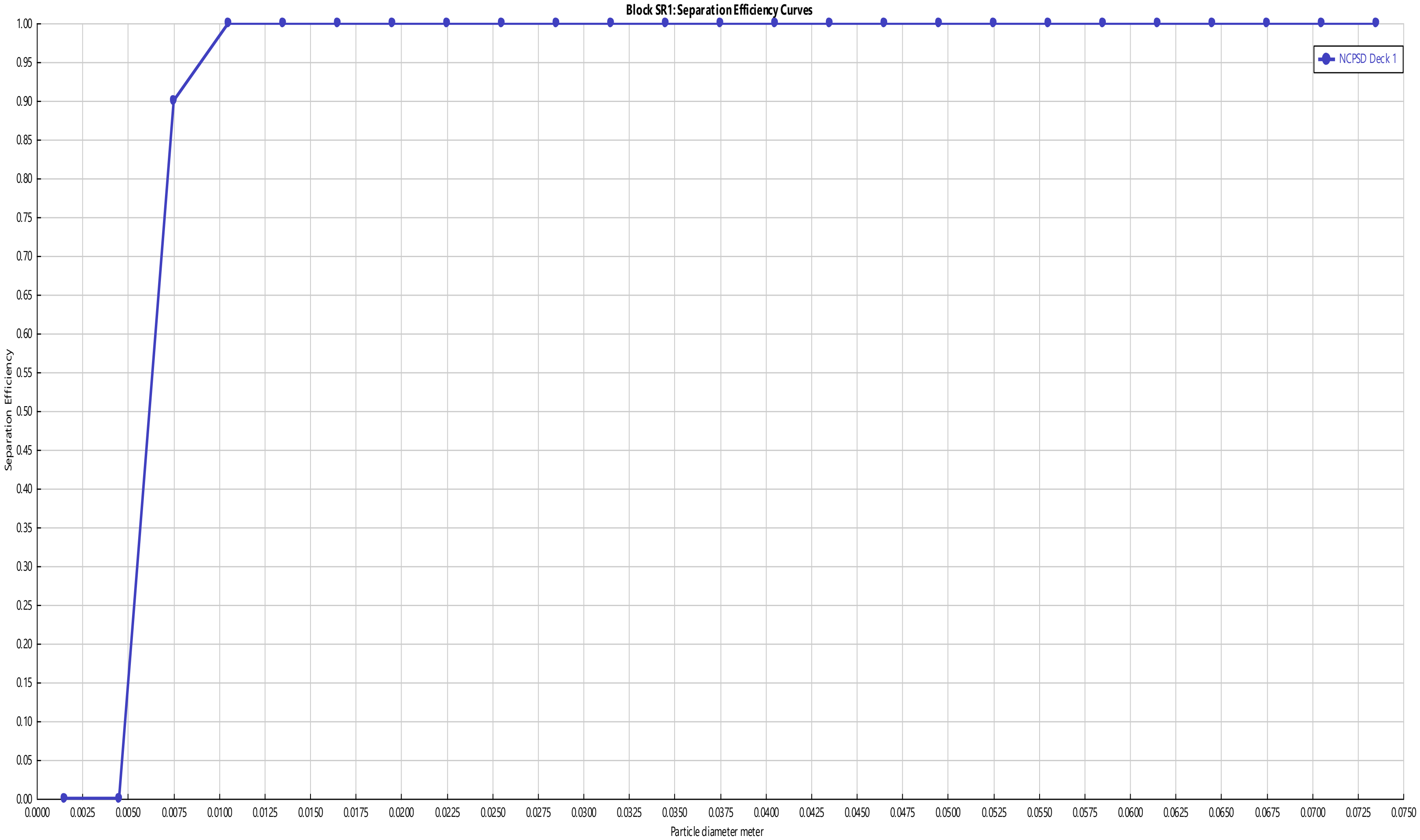
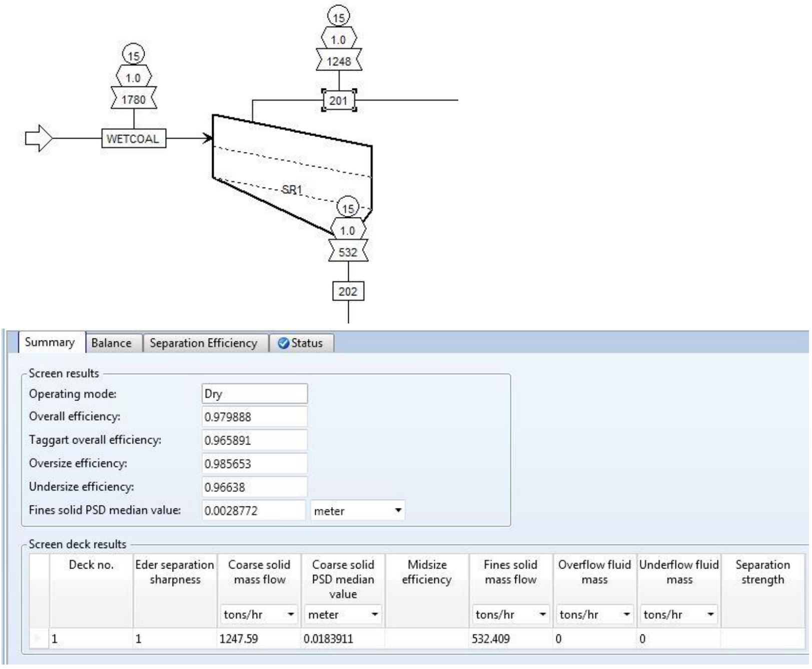


Fig 20: Primary screening separation and efficiency

The details of the primary SCREENING machine is presented in **Fig 21.**



Screen results

Operating mode: Dry
Overall efficiency: 0.979888
Taggart overall efficiency: 0.965891
Oversize efficiency: 0.985653
Undersize efficiency: 0.96638
Fines solid PSD median value: 0.0028772 meter

Screen deck results

Deck no.	Eder separation sharpness	Coarse solid mass flow	Coarse solid PSD median value	Midsize efficiency	Fines solid mass flow	Overflow fluid mass	Underflow fluid mass	Separation strength
		tons/hr	meter		tons/hr	tons/hr	tons/hr	
1	1	1247.59	0.0183911		532.409	0	0	

Fig 21: Details of primary screener

The hammer mill is considered as a HAMMER CRUSHER with default inbuilt model. The PSD curve shape, i.e. RRSB distribution function using dispersion parameter=2, D63 particle size adjusted to suit ~98 wt% particles <9.3 mm and secondary split. The open cut method with Bands law (HG-index 130) is used.

The PSD after primary hammering is presented in **Fig 22.**

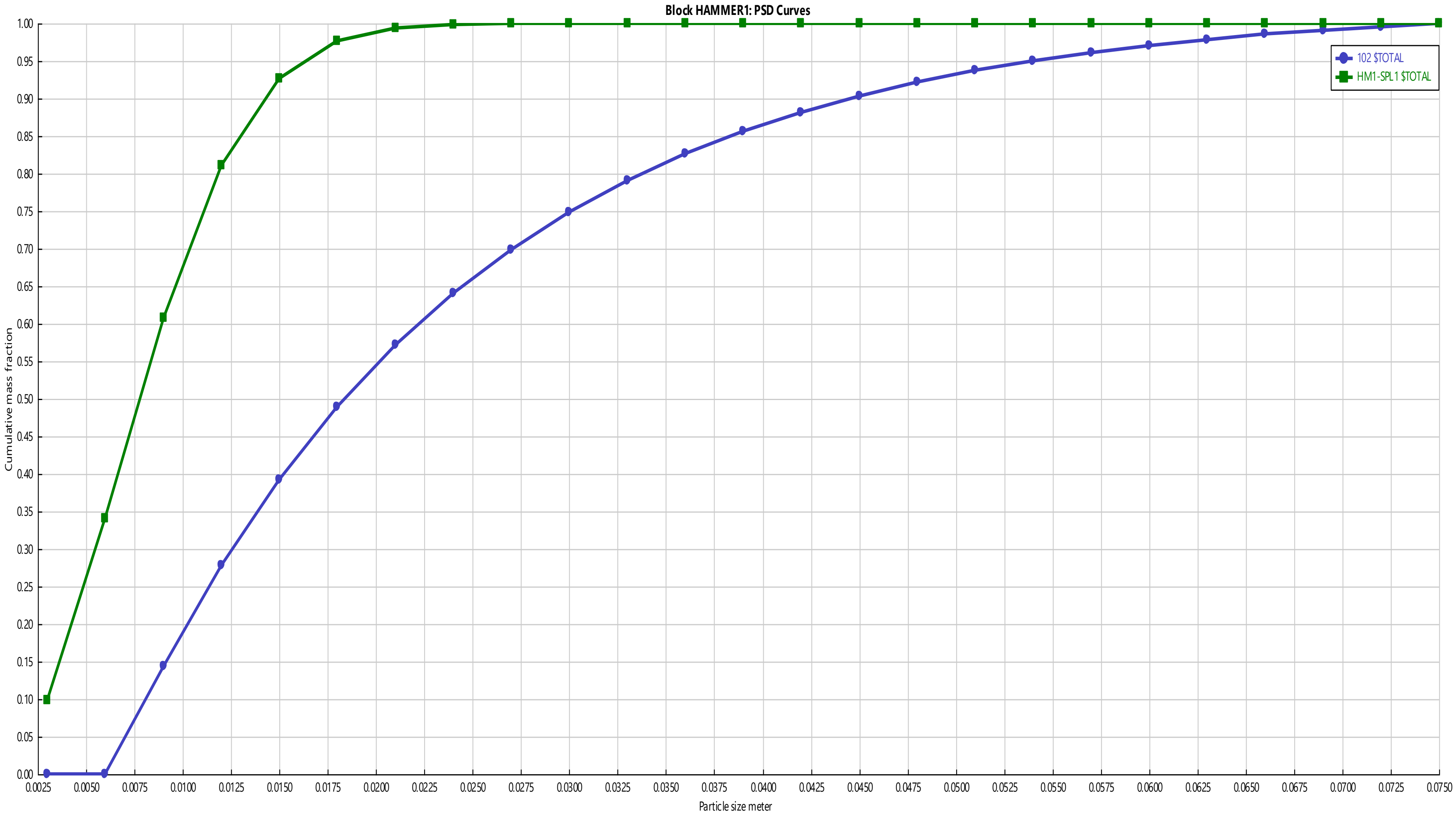
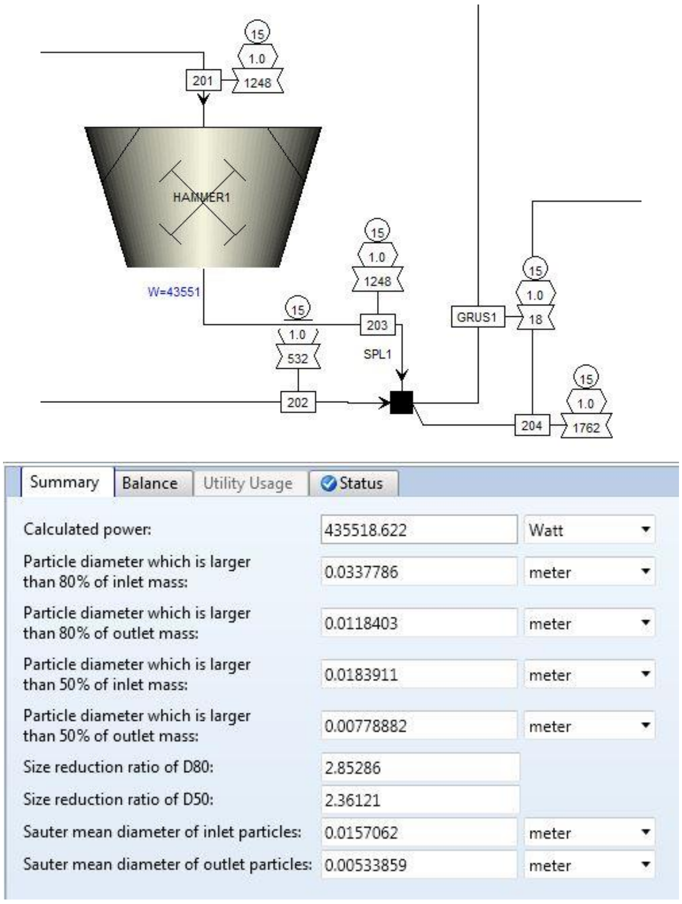


Fig 22: PSD after primary hammering

The details of the primary Hammer machine is presented in **Fig 23.**



Calculated power: 435518.622 Watt

Particle diameter which is larger than 80% of inlet mass: 0.0337786 meter

Particle diameter which is larger than 80% of outlet mass: 0.0118403 meter

Particle diameter which is larger than 50% of inlet mass: 0.0183911 meter

Particle diameter which is larger than 50% of outlet mass: 0.00778882 meter

Size reduction ratio of D80: 2.85286

Size reduction ratio of D50: 2.36121

Sauter mean diameter of inlet particles: 0.0157062 meter

Sauter mean diameter of outlet particles: 0.00533859 meter

Fig 23: Details of Primary Hammer

**1% of raw coal was considered as grus and was not accounted in the mainstream. An FSPLIT was used for this purpose.

Steam Tube dryer

The steam tube dryer is modelled with 3 different default inbuilt unit operation unit. An RStoic is used to convert the non-conventional moisture present in coal to convert conventional pure water at 15°C based on Eq (1):

$$\text{WETCOAL} \rightarrow 0.0555084\ H_2O \text{ ……..} (1)$$

The pure conventional water is passed through a heat exchanger (HX) with a specification of vapour fraction 1 at the outlet of the cold fluid. All the water that is vaporised in the HX unit is separate in a splitter (SP1). **1% of entrainment of coal was considered in the water vapour.** The overall unit of the dryer is presented in **Fig 24.**

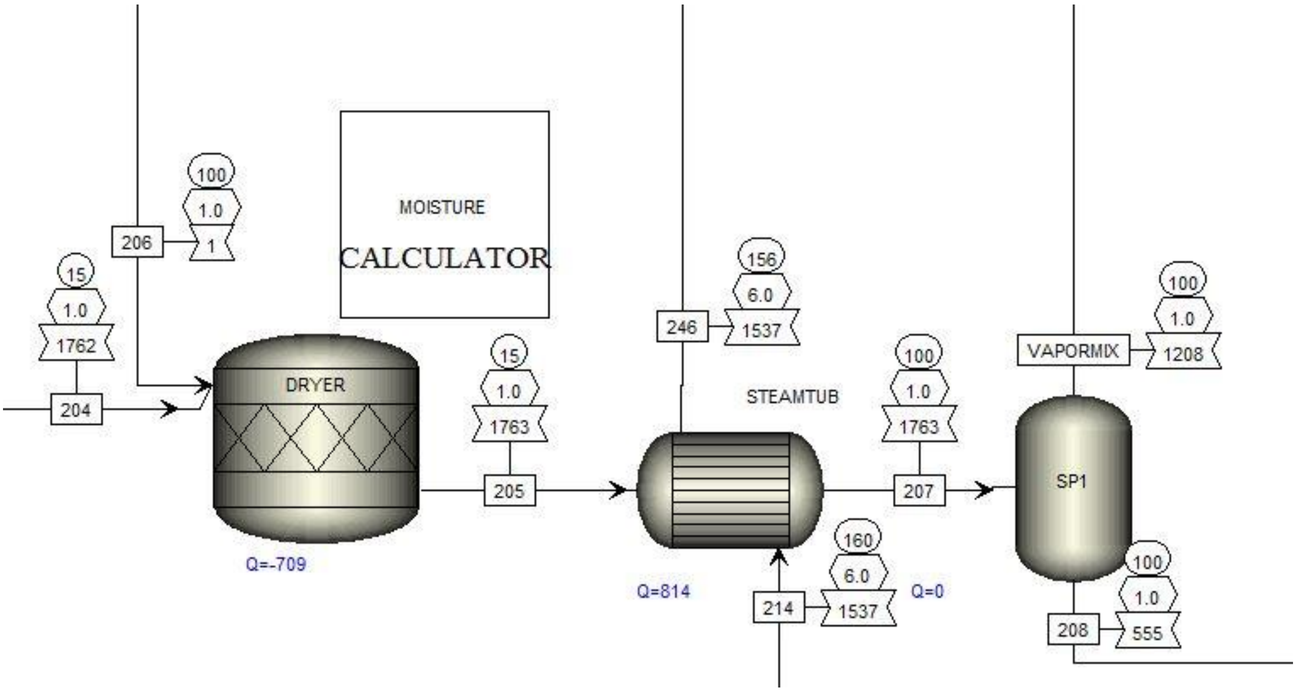
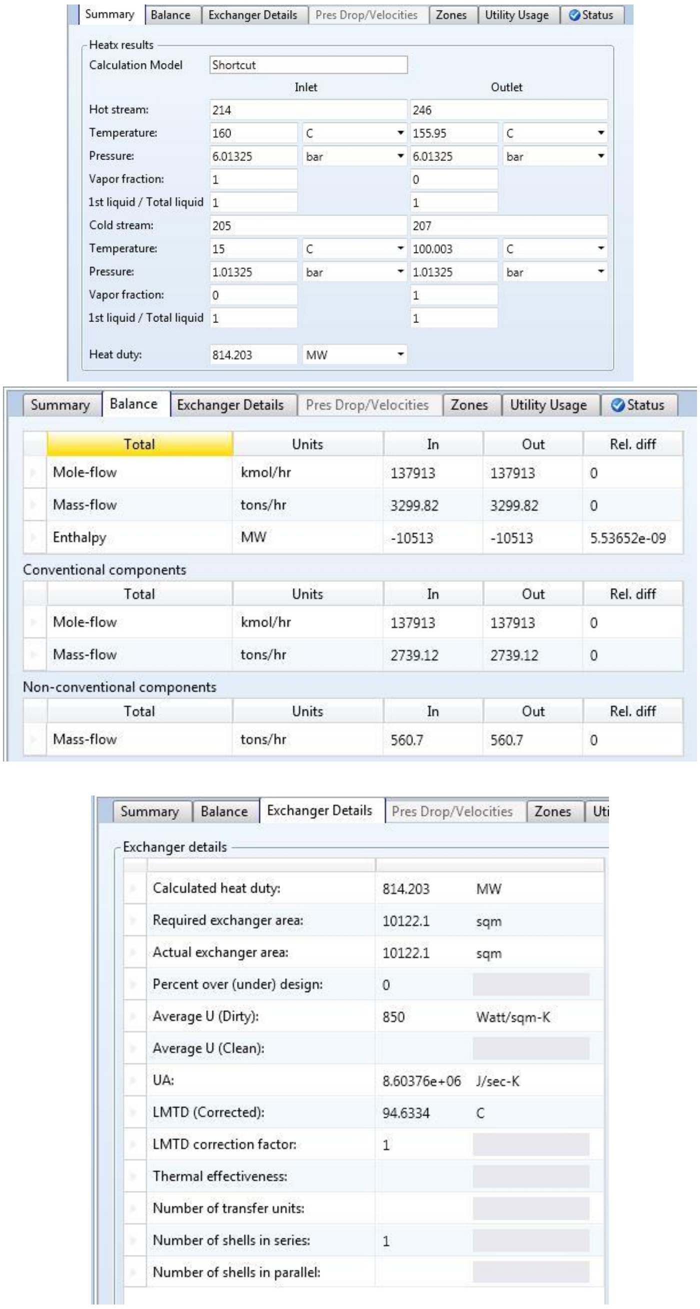


Fig 24: Steam tube dryer

A stream of air 206 is introduced in the DRYER to pull out the moisture from the tubes. 160°C and 5 barg steam were used in the hot stream of the HX. The steam requirement was adjusted such a way that the temperature drop after HX of the hot stream is 156°C.The details of the HX unit is presented in **Fig 25.**



Heatx results

	Inlet		Outlet	
Calculation Model	Shortcut			
Hot stream:	214		246	
Temperature:	160	C	155.95	C
Pressure:	6.01325	bar	6.01325	bar
Vapor fraction:	1		0	
1st liquid / Total liquid	1		1	
Cold stream:	205		207	
Temperature:	15	C	100.003	C
Pressure:	1.01325	bar	1.01325	bar
Vapor fraction:	0		1	
1st liquid / Total liquid	1		1	
Heat duty:	814.203	MW		

Total	Units	In	Out	Rel. diff
Mole-flow	kmol/hr	137913	137913	0
Mass-flow	tons/hr	3299.82	3299.82	0
Enthalpy	MW	-10513	-10513	5.53652e-09

Conventional components

Total	Units	In	Out	Rel. diff
Mole-flow	kmol/hr	137913	137913	0
Mass-flow	tons/hr	2739.12	2739.12	0

Non-conventional components

Total	Units	In	Out	Rel. diff
Mass-flow	tons/hr	560.7	560.7	0

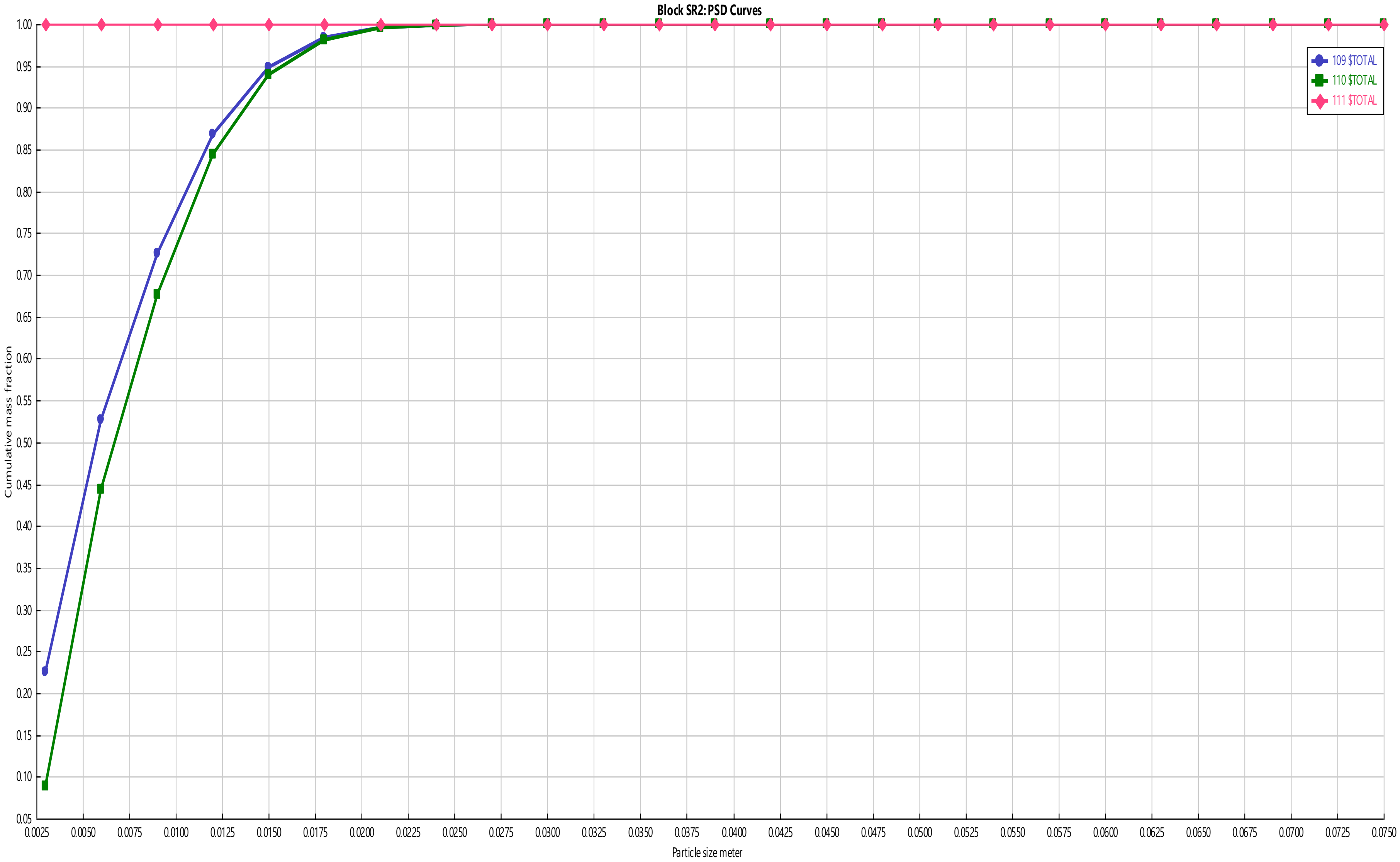
Exchanger details

Calculated heat duty:	814.203	MW
Required exchanger area:	10122.1	sqm
Actual exchanger area:	10122.1	sqm
Percent over (under) design:	0	
Average U (Dirty):	850	Watt/sqm-K
Average U (Clean):		
UA:	8.60376e+06	J/sec-K
LMTD (Corrected):	94.6334	C
LMTD correction factor:	1	
Thermal effectiveness:		
Number of transfer units:		
Number of shells in series:	1	
Number of shells in parallel:		

Fig 25: Details of Steam tube dryer unit

Secondary screening and Hammering

The dry coal goes to a secondary screening and hammering for further refinement. The secondary screener is modelled as default SCREEN with a cut size of 0.002 m (Coarse-split) to meet the secondary Hammer mill requirement. The separation stage and efficiency is presented in **Fig 26**.


Block SR2: PSD Curves
1.00
0.95
0.90
0.85
0.80
0.75
0.70
0.65
0.60
0.55
0.50
0.45
0.40
0.35
0.30
0.25
0.20
0.15
0.10
0.05
Cumulative mass fraction
0.0025
0.0050
0.0075
0.0100
0.0125
0.0150
0.0175
0.0200
0.0225
0.0250
0.0275
0.0300
0.0325
0.0350
0.0375
0.0400
0.0425
0.0450
0.0475
0.0500
0.0525
0.0550
0.0575
0.0600
0.0625
0.0650
0.0675
0.0700
0.0725
0.0750
Particle size meter
109 $TOTAL
110 $TOTAL
111 $TOTAL

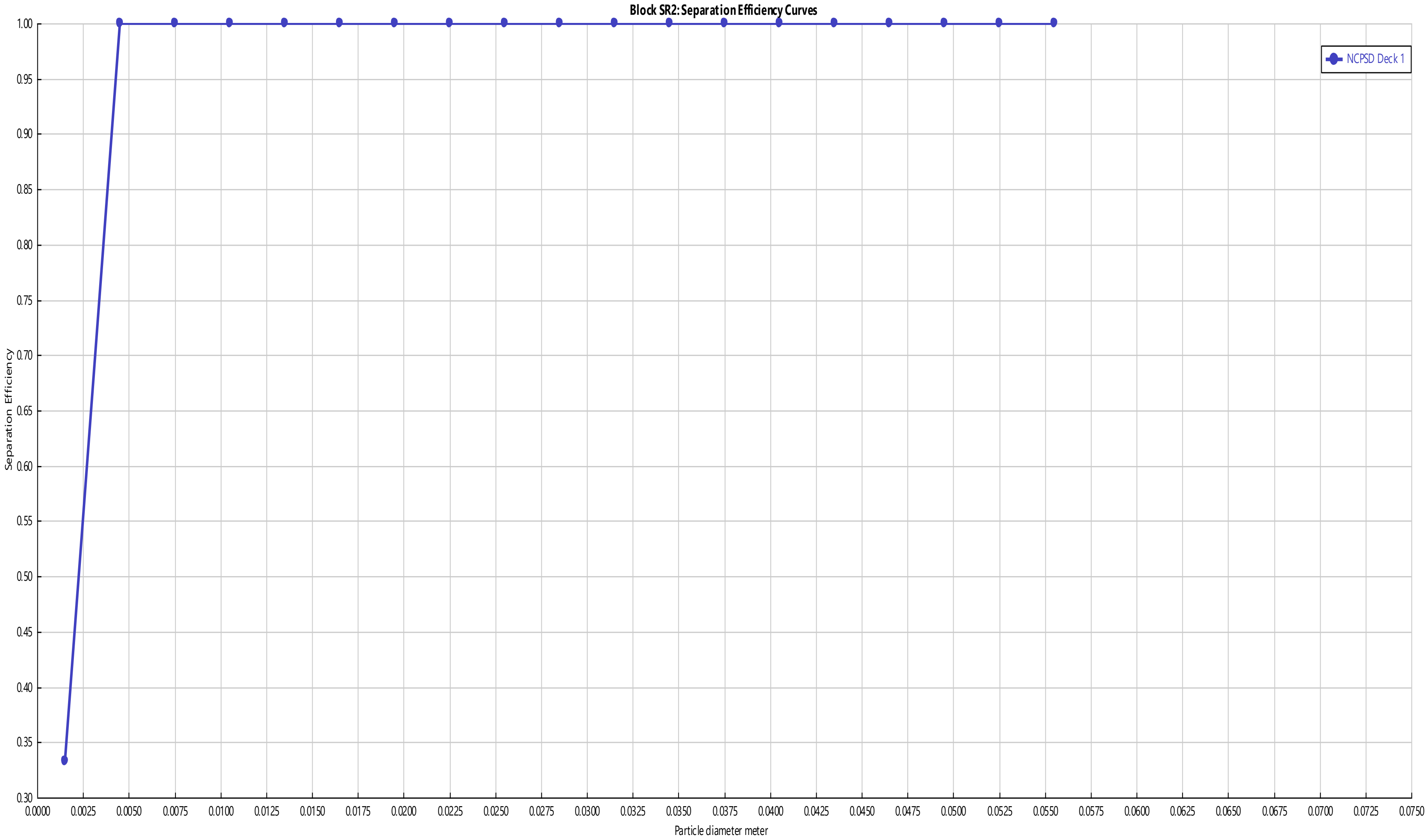
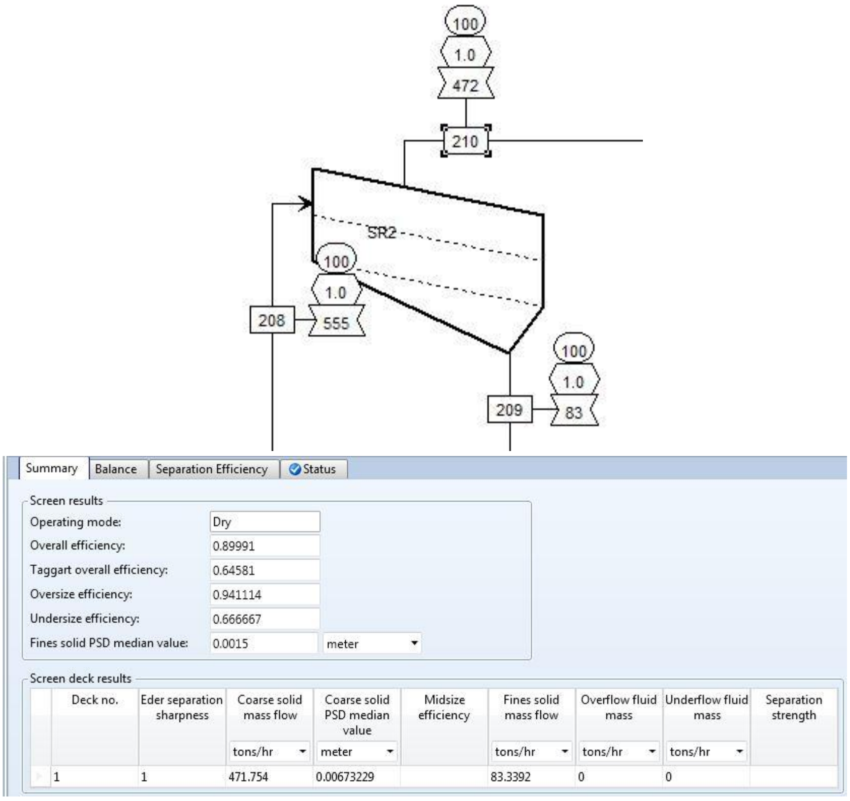

Block SR2: Separation Efficiency Curves
NCPSD Deck 1
Separation Efficiency
1.00
0.95
0.90
0.85
0.80
0.75
0.70
0.65
0.60
0.55
0.50
0.45
0.40
0.35
0.30
0.0000
0.0025
0.0050
0.0075
0.0100
0.0125
0.0150
0.0175
0.0200
0.0225
0.0250
0.0275
0.0300
0.0325
0.0350
0.0375
0.0400
0.0425
0.0450
0.0475
0.0500
0.0525
0.0550
0.0575
0.0600
0.0625
0.0650
0.0675
0.0700
0.0725
0.0750
Particle diameter meter

Fig 26: Secondary screening separation and efficiency

The details of the primary SCREENING machine is presented in **Fig 27**



Screen results

Operating mode: Dry
Overall efficiency: 0.89991
Taggart overall efficiency: 0.64581
Oversize efficiency: 0.941114
Undersize efficiency: 0.666667
Fines solid PSD median value: 0.0015 meter

Screen deck results

Deck no.	Eder separation sharpness	Coarse solid mass flow tons/hr	Coarse solid PSD median value meter	Midsize efficiency	Fines solid mass flow tons/hr	Overflow fluid mass tons/hr	Underflow fluid mass tons/hr	Separation strength
1	1	471.754	0.00673229		83.3392	0	0	

Fig 27: Details of secondary screener

The secondary hammer mill is considered as a HAMMER CRUSHER with default inbuilt model. The PSD curve shape, i.e. RRSB distribution function using dispersion parameter=2, D63 particle size adjusted to suit ~98 wt% particles <2 mm and secondary split.

The PSD after primary hammering is presented in **Fig 28.**

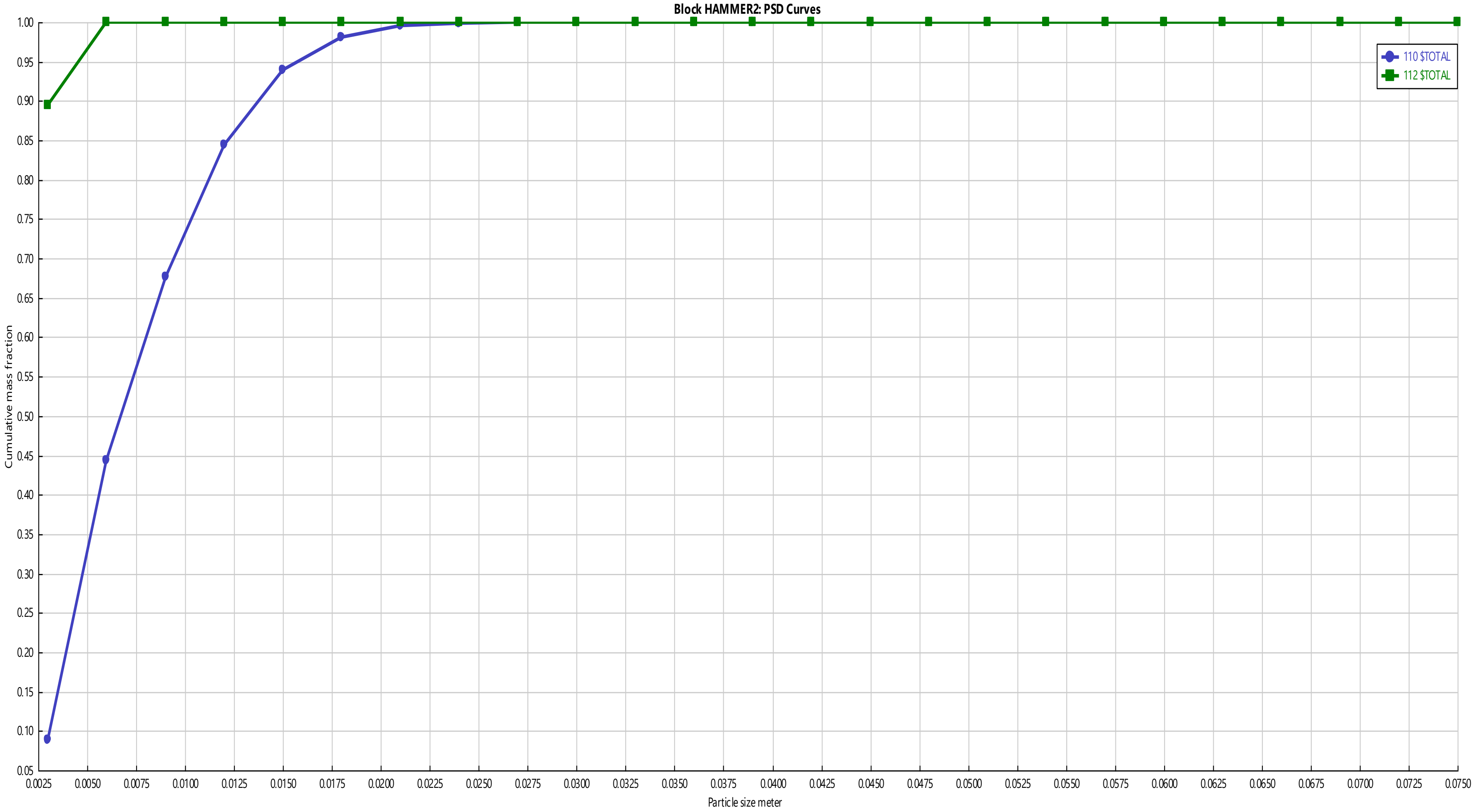
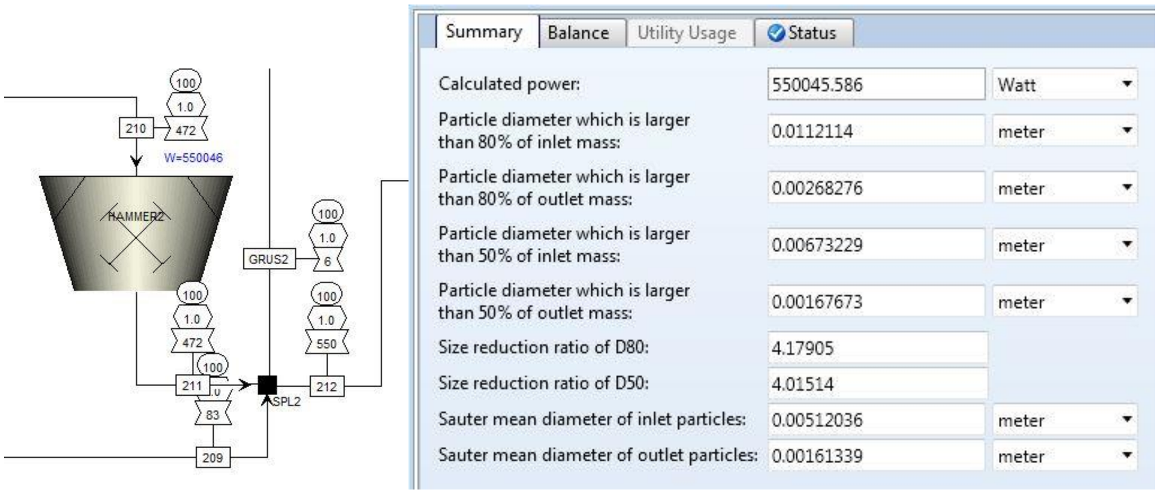


Fig 28: PSD of Secondary Hammer

The details of the secondary Hammer machine is presented in **Fig 29.**



Calculated power:	550045.586	Watt
Particle diameter which is larger than 80% of inlet mass:	0.0112114	meter
Particle diameter which is larger than 80% of outlet mass:	0.00268276	meter
Particle diameter which is larger than 50% of inlet mass:	0.00673229	meter
Particle diameter which is larger than 50% of outlet mass:	0.00167673	meter
Size reduction ratio of D80:	4.17905	
Size reduction ratio of D50:	4.01514	
Sauter mean diameter of inlet particles:	0.00512036	meter
Sauter mean diameter of outlet particles:	0.00161339	meter

Fig 29: Details of Secondary Hammer

**1% of raw coal was considered as grus and was not accounted in the mainstream. An FSPLIT was used for this purpose.

Briquette unit

The briquette unit is modelled as default granulator model of process simulation package. The briquette is considered as 62 mm sphere diameter for an ellipsoid containing a dimension of 79x80x30 equivalent to 126 cm^3. The PSD is shown in **Fig 30** and the details in **Fig 31.**

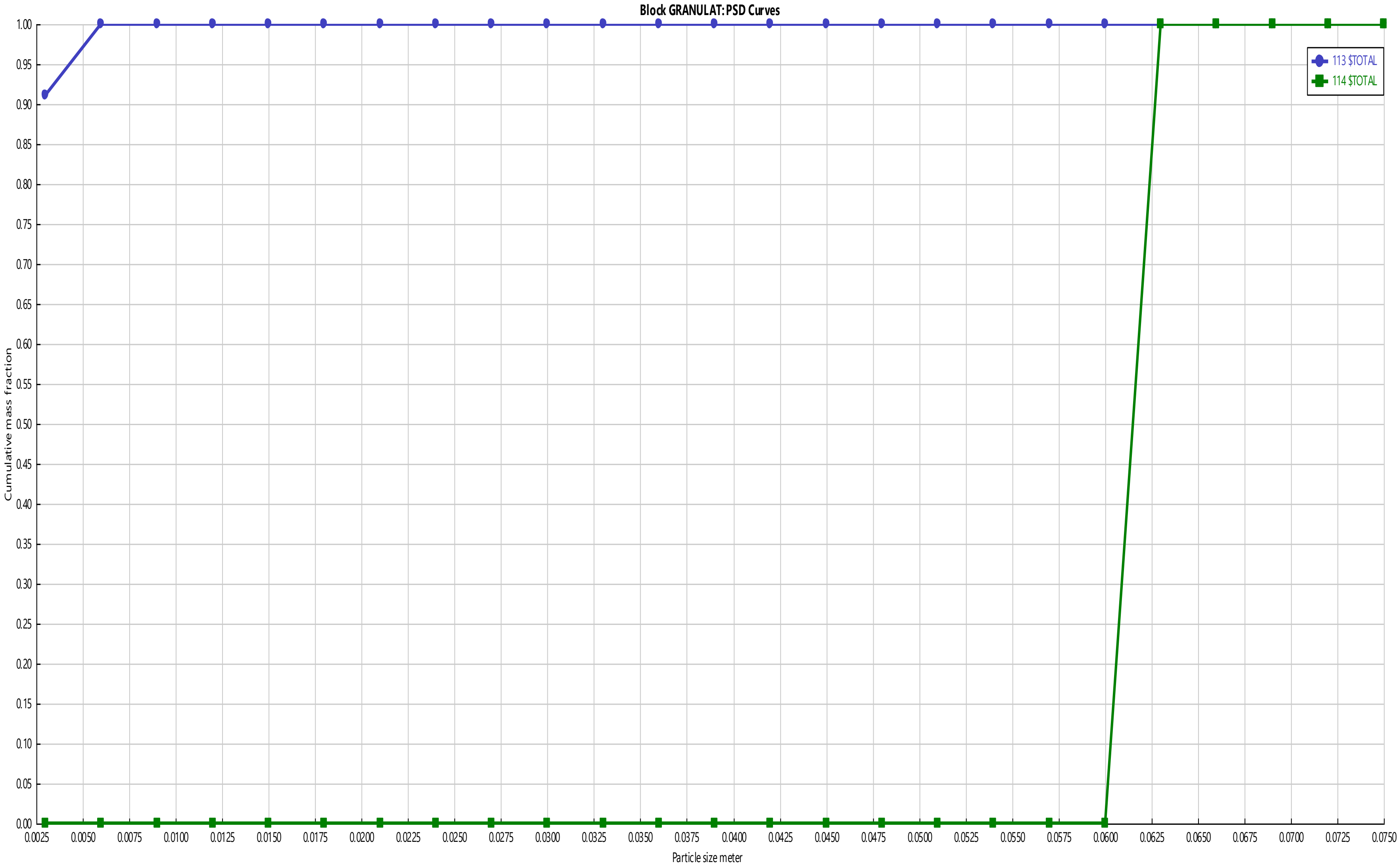
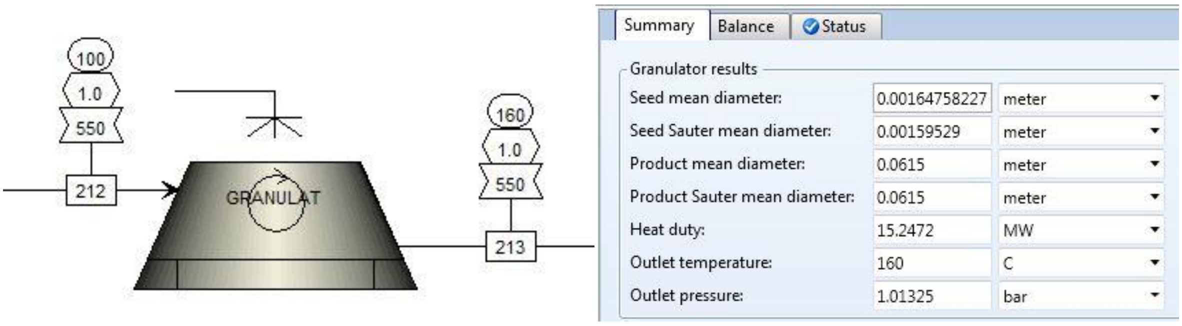


Fig 30: PSD in Granulator



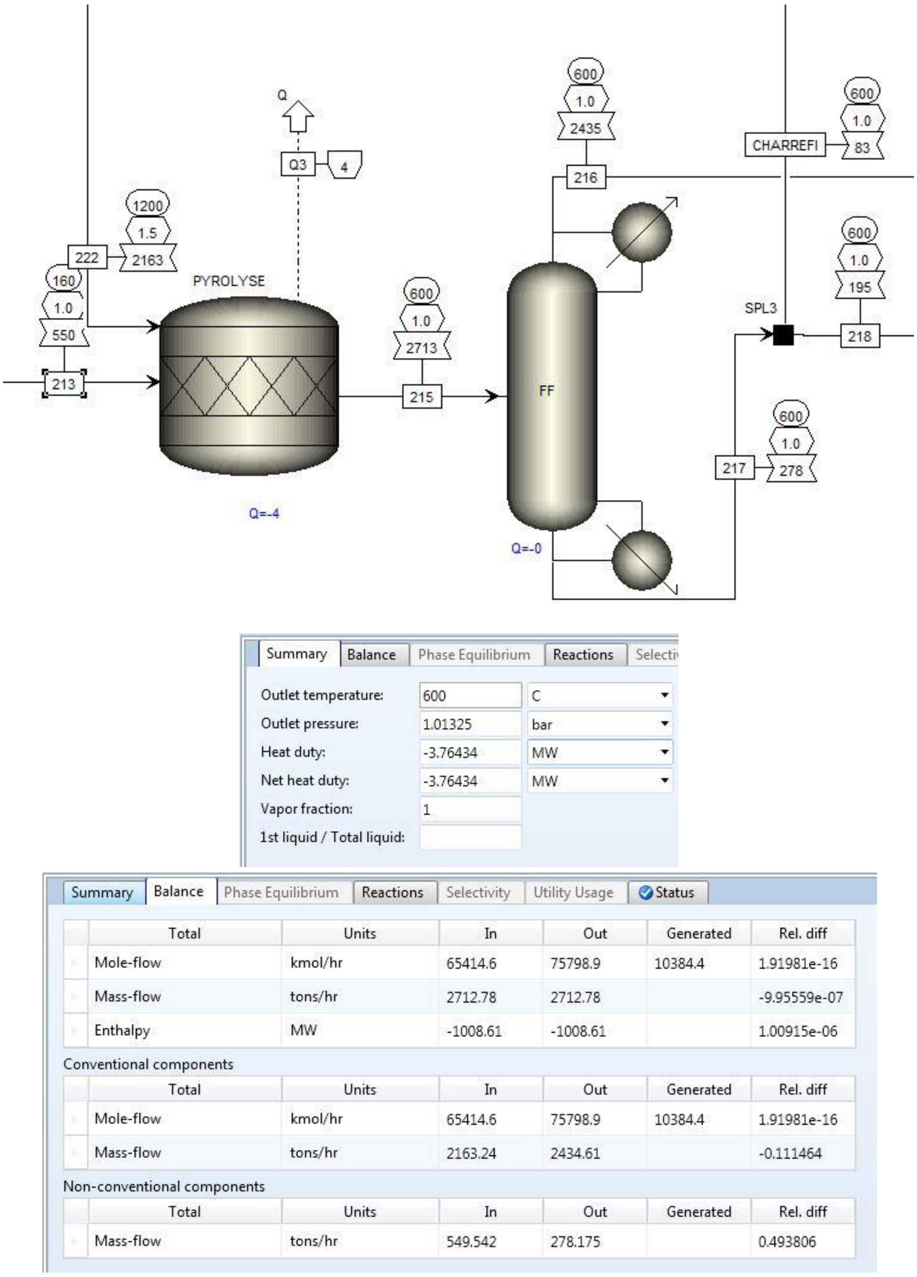
Granulator results

Seed mean diameter: 0.00164758227 meter
Seed Sauter mean diameter: 0.00159529 meter
Product mean diameter: 0.0615 meter
Product Sauter mean diameter: 0.0615 meter
Heat duty: 15.2472 MW
Outlet temperature: 160 C
Outlet pressure: 1.01325 bar

Fig 31: Details of Granulator unit

Pyrolyser unit

The Pyrolyser unit is modelled as an RStoic reactor with a specific reactor temperature of 600°C and a component splitter. The heat requirement for the unit is satisfied by direct passing hot flue gas from the combustor. We assumed the flue gases from the combustor do not react with the pyrolysis products. The heat requirement is adjusted through the extent of char burning. It was found 70% of the char should be burnt to meet the energy requirement of the pyrolyzer. **Fig 32** present the pyrolyzer unit.



Outlet temperature: 600 C
Outlet pressure: 1.01325 bar
Heat duty: -3.76434 MW
Net heat duty: -3.76434 MW
Vapor fraction: 1
1st liquid / Total liquid:

Total	Units	In	Out	Generated	Rel. diff
Mole-flow	kmol/hr	65414.6	75798.9	10384.4	1.91981e-16
Mass-flow	tons/hr	2712.78	2712.78		-9.95559e-07
Enthalpy	MW	-1008.61	-1008.61		1.00915e-06

Conventional components

Total	Units	In	Out	Generated	Rel. diff
Mole-flow	kmol/hr	65414.6	75798.9	10384.4	1.91981e-16
Mass-flow	tons/hr	2163.24	2434.61		-0.111464

Non-conventional components

Total	Units	In	Out	Generated	Rel. diff
Mass-flow	tons/hr	549.542	278.175		0.493806

Fig 32: Details of the pyrolyzer unit

Combustor unit

The combustor unit was modelled as an RStoic, RGibbs and a Cyclone separator presented in **Fig 33**. The RStoic is used to convert the char into its preliminary constitution. Standard combustion reactions were considered in the RGibbs reactor. The Ash was separated by Cyclone separator.

The combustor operates at 1200°C, 1.5 bar pressure. The air requirement for the char combustion was calculated with a calculator block. The air was preheated to 250°C through the hot pyrolysis gases by a heat exchanger (HX1) with a specific cold outlet temperature of 250°C. 70% of the char from the pyrolysis unit was combusted to meet the energy requirement of the pyrolyzer and other accessories.

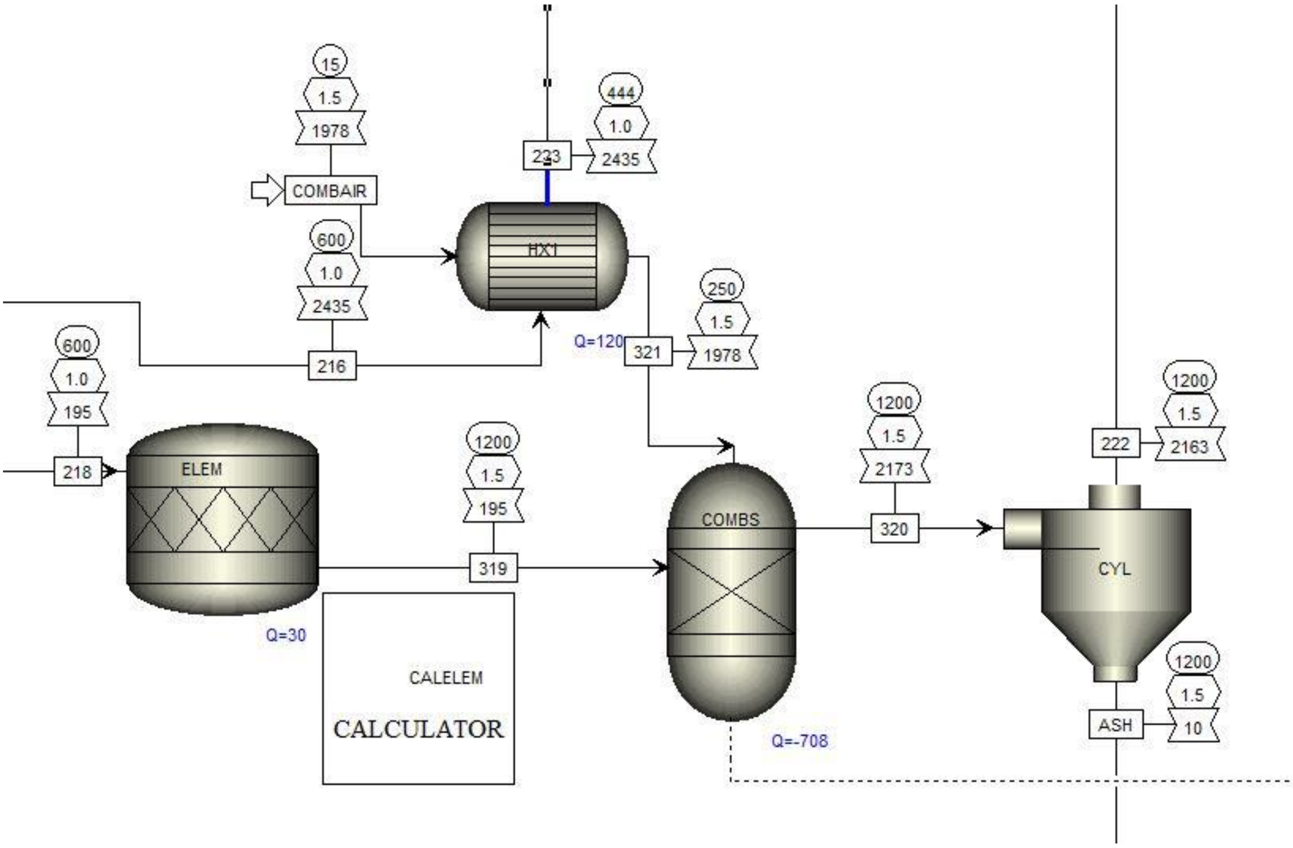
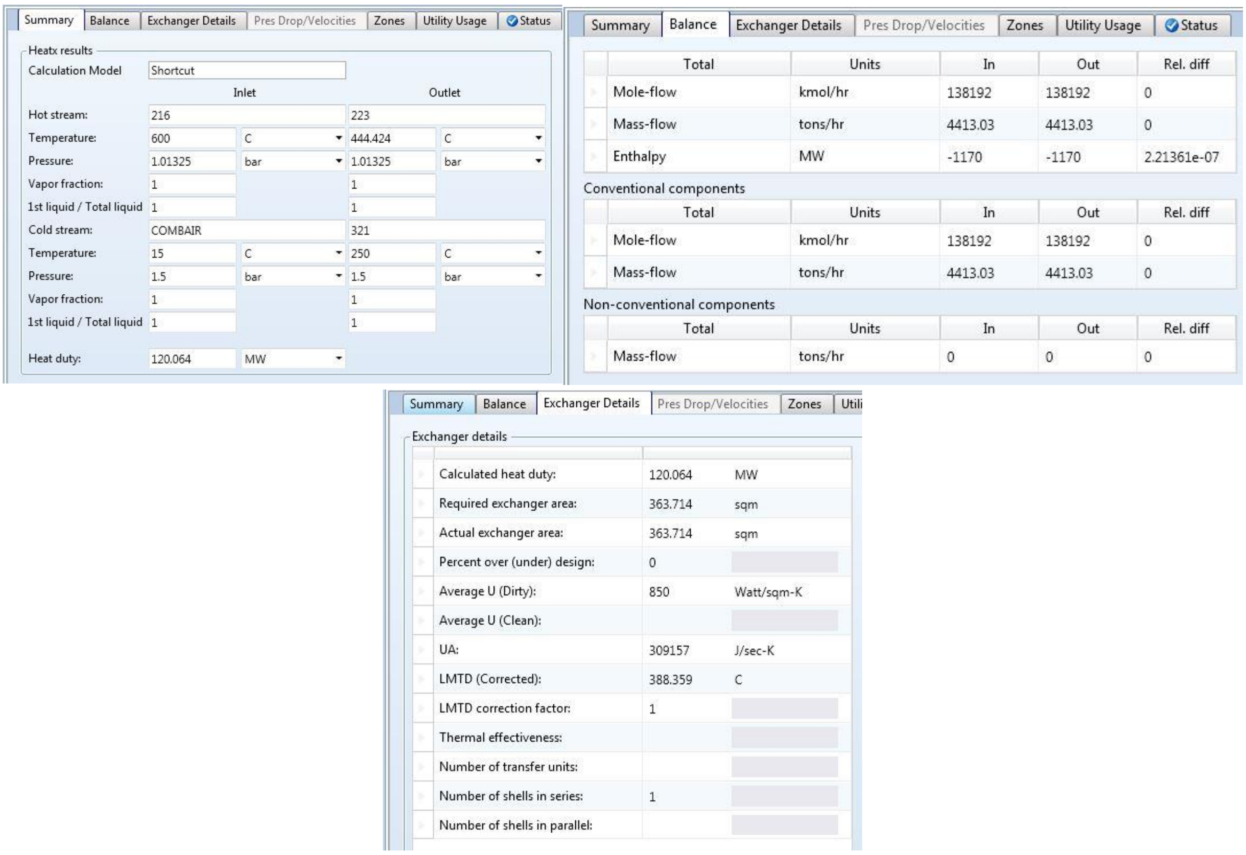


Fig 33: Combustor unit

The details of the HX1 unit is presented in **Fig 34.**



Heatx results

Calculation Model	Shortcut			
	Inlet		Outlet	
Hot stream:	216		223	
Temperature:	600	C	444.424	C
Pressure:	1.01325	bar	1.01325	bar
Vapor fraction:	1		1	
1st liquid / Total liquid	1		1	
Cold stream:	COMBAIR		321	
Temperature:	15	C	250	C
Pressure:	1.5	bar	1.5	bar
Vapor fraction:	1		1	
1st liquid / Total liquid	1		1	
Heat duty:	120.064	MW		

Total	Units	In	Out	Rel. diff
Mole-flow	kmol/hr	138192	138192	0
Mass-flow	tons/hr	4413.03	4413.03	0
Enthalpy	MW	-1170	-1170	2.21361e-07

Conventional components

Total	Units	In	Out	Rel. diff
Mole-flow	kmol/hr	138192	138192	0
Mass-flow	tons/hr	4413.03	4413.03	0

Non-conventional components

Total	Units	In	Out	Rel. diff
Mass-flow	tons/hr	0	0	0

Exchanger details

Calculated heat duty:	120.064	MW
Required exchanger area:	363.714	sqm
Actual exchanger area:	363.714	sqm
Percent over (under) design:	0	
Average U (Dirty):	850	Watt/sqm-K
Average U (Clean):		
UA:	309157	J/sec-K
LMTD (Corrected):	388.359	C
LMTD correction factor:	1	
Thermal effectiveness:		
Number of transfer units:		
Number of shells in series:	1	
Number of shells in parallel:		

Fig 34: Details of air preheater for char combustor

Pyrolysis Oil extraction unit

The stream of pyrolysis gas and the rest 20% flue gas were further cooled down to 25°C with a simple Heater to calculate the energy available to produce steam for the dryer unit. The liquid-vapour was separated with a FLASH unit. **Fig 35** presents the scheme.

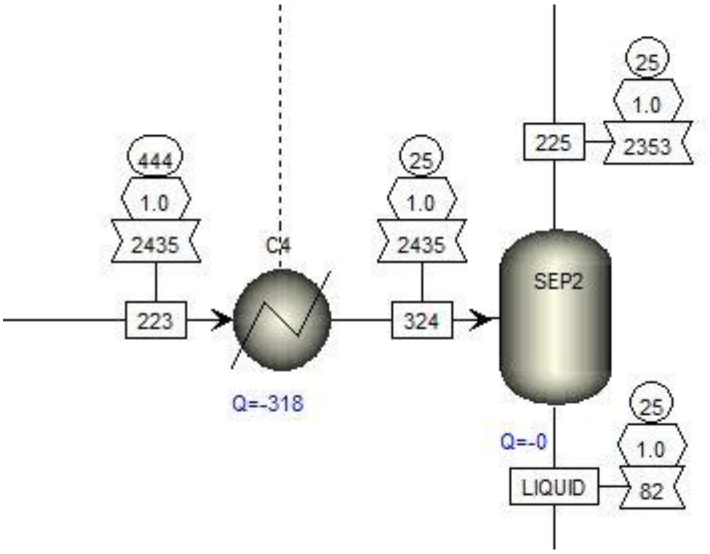
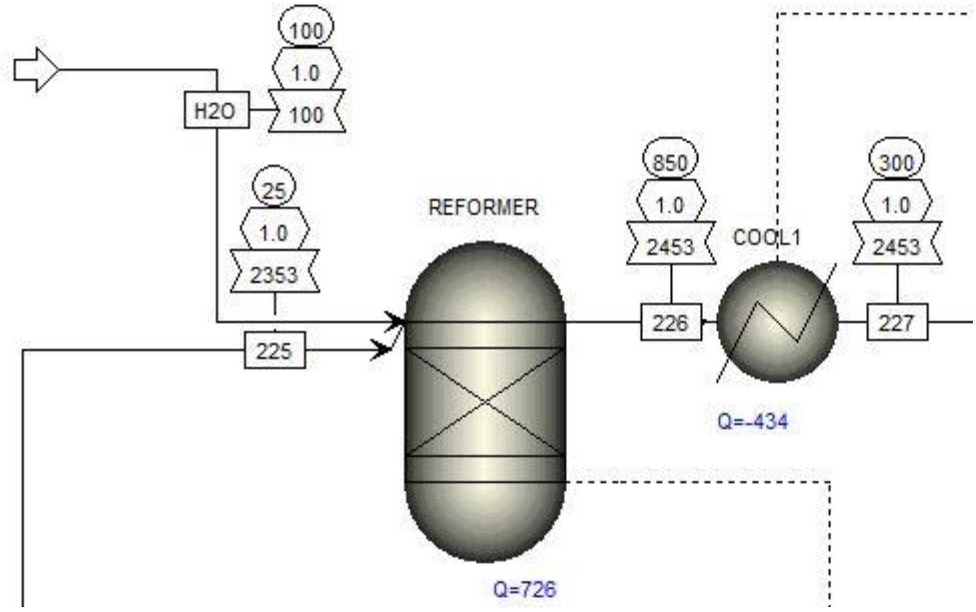
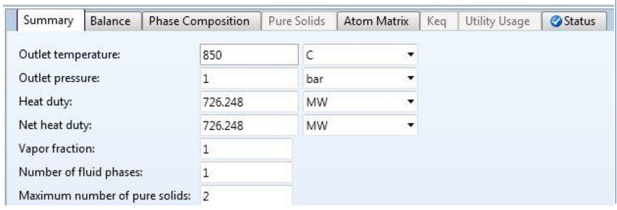


Fig 35: Pyrolysis Oil extraction unit

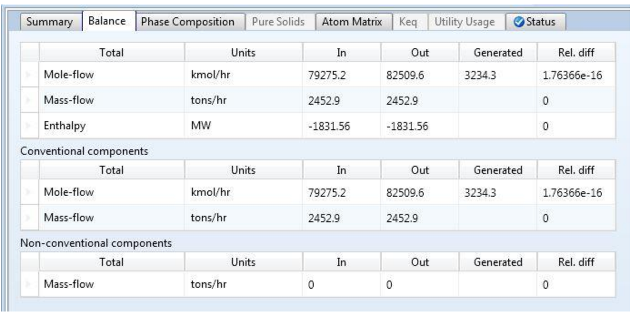
Reformer and heat recovery unit

The Reformer unit is modelled as an RGibbs reactor with a specified reactor temperature of 850°C. The heat requirement for this reactor is satisfied from the heat energy coming from the combustor unit. The outlet gases from the reactor are cooled at 300°C by a HEATER block (Cooler 1). The energy from this block is used to heat the gases after SHIFT reactor to meet the energy requirements of CALCINATOR reactor. The details of this unit are presented in **Fig 36.** The steam requirement for this adjusted in such a way so that most of the Metahne and Ethane is converted into CO and H_2.





Outlet temperature: 850 C
Outlet pressure: 1 bar
Heat duty: 726.248 MW
Net heat duty: 726.248 MW
Vapor fraction: 1
Number of fluid phases: 1
Maximum number of pure solids: 2



Total	Units	In	Out	Generated	Rel. diff
Mole-flow	kmol/hr	79275.2	82509.6	3234.3	1.76366e-16
Mass-flow	tons/hr	2452.9	2452.9		0
Enthalpy	MW	-1831.56	-1831.56		0

Conventional components

Total	Units	In	Out	Generated	Rel. diff
Mole-flow	kmol/hr	79275.2	82509.6	3234.3	1.76366e-16
Mass-flow	tons/hr	2452.9	2452.9		0

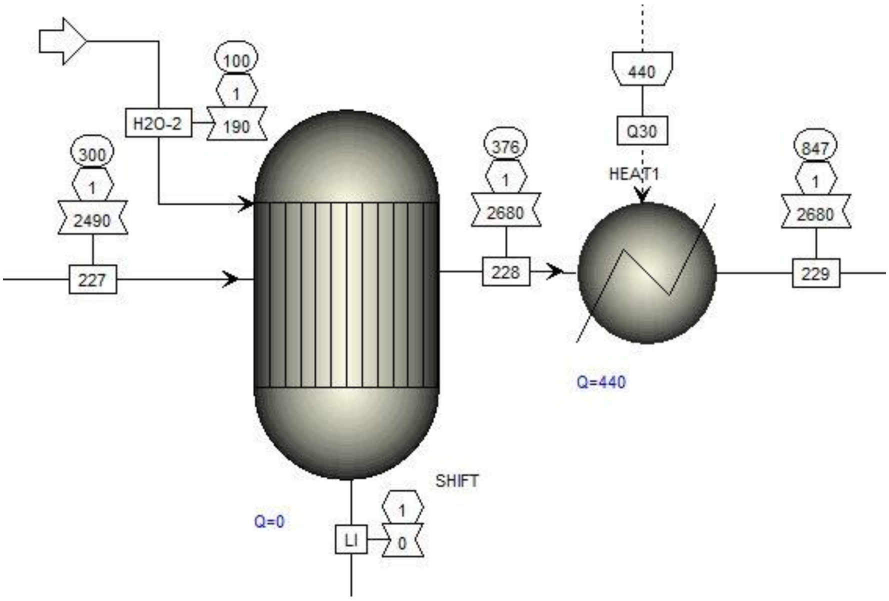
Non-conventional components

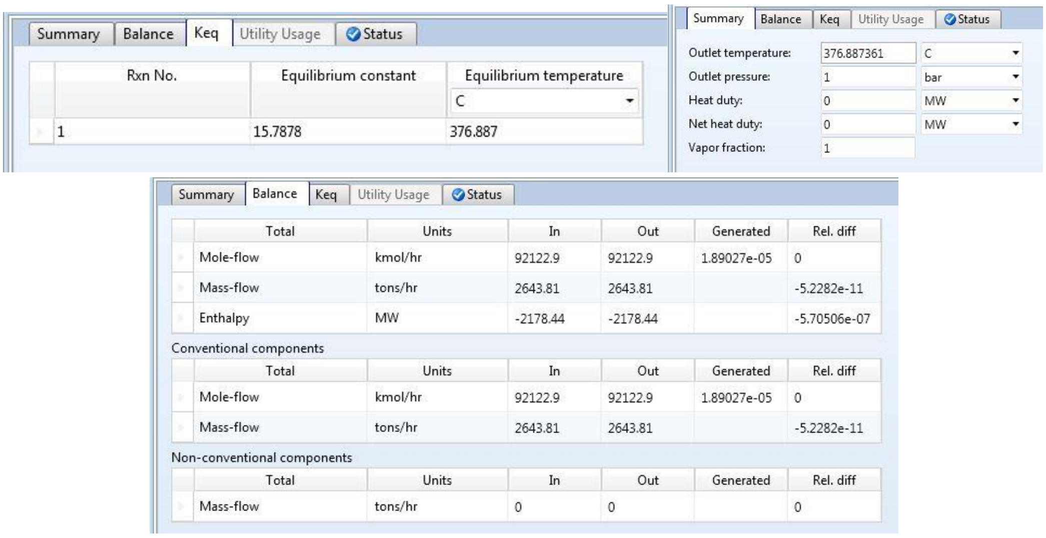
Total	Units	In	Out	Generated	Rel. diff
Mass-flow	tons/hr	0	0		0

Fig 36: Details of Reformer unit

SHIFT reactor and heat recovery unit

The SHIFT reactor is modelled as a REquil reactor. The steam requirement for this unit is calculated through a calculator block.





Rxn No.	Equilibrium constant	Equilibrium temperature C
1	15.7878	376.887

Outlet temperature:	376.887361	C
Outlet pressure:	1	bar
Heat duty:	0	MW
Net heat duty:	0	MW
Vapor fraction:	1	

Total	Units	In	Out	Generated	Rel. diff
Mole-flow	kmol/hr	92122.9	92122.9	1.89027e-05	0
Mass-flow	tons/hr	2643.81	2643.81		-5.2282e-11
Enthalpy	MW	-2178.44	-2178.44		-5.70506e-07

Conventional components

Total	Units	In	Out	Generated	Rel. diff
Mole-flow	kmol/hr	92122.9	92122.9	1.89027e-05	0
Mass-flow	tons/hr	2643.81	2643.81		-5.2282e-11

Non-conventional components

Total	Units	In	Out	Generated	Rel. diff
Mass-flow	tons/hr	0	0		0

Fig 37: Details of Shift reactor unit

Gas cleaning and CO_2 separation unit

The gas cleaning and CO_2 separation unit was modelled as a $CACO_3$ looping unit presented in **Fig 38**. An extensive study was carried out, and we choose this option. $CACO_3$ looping can meet dual purposes, the pollutant removal as well as the separation of CO_2 from the mainstream. The main advantage of this unit is it does not require fresh water stream as like Ammine scrubbing. There is two RGibbs reactors with two cyclone separator. In the carbonator reactor, the reaction enthalpy is exothermic and specified as 650°C. The stream of gases 229 enters the reactor and reacts with CAO.

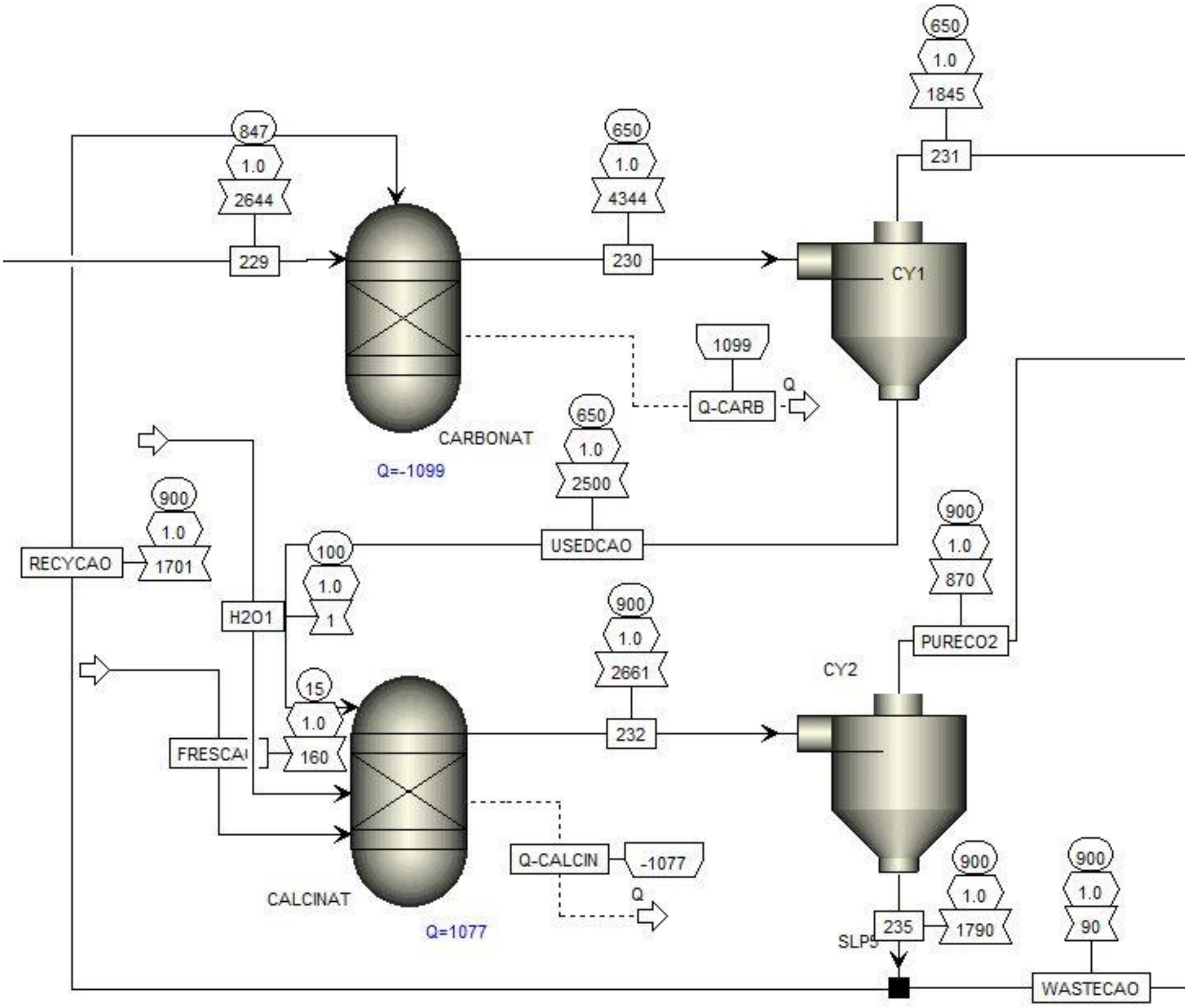


Fig 38: $CaCO_3$ looping unit for gas cleaning and CO_2 separation

The reaction enthalpy in the CALCINATOR is endothermic. The heat requirement of this block is satisfied by the excess heat energy from the CARBONATOR unit. The CARBONATOR block will have excess energy to run the CALCINATOR if efficient heat removal is done before the REFORMER reactor that is used to heat the gases before CARBONATOR.

5% of the CAO was considered as waste. The entire unit was modelled in such a way so that in the recycle stream (USED CAO) there is no CAO going to CALCINATOR unit. A flow of steam (H2O) was added to the CALCINATOR so that it can drive the CO_2 and other pollutant gases from the system.

Heat recovery and PSA for H2 Separation

Energy is recovered from the clean gas and pure CO_2 stream through HEATER block as presented in **Fig 39**. The energy is then used for steam generation for the dryer. The H2 separation unit is considered as a component splitter.

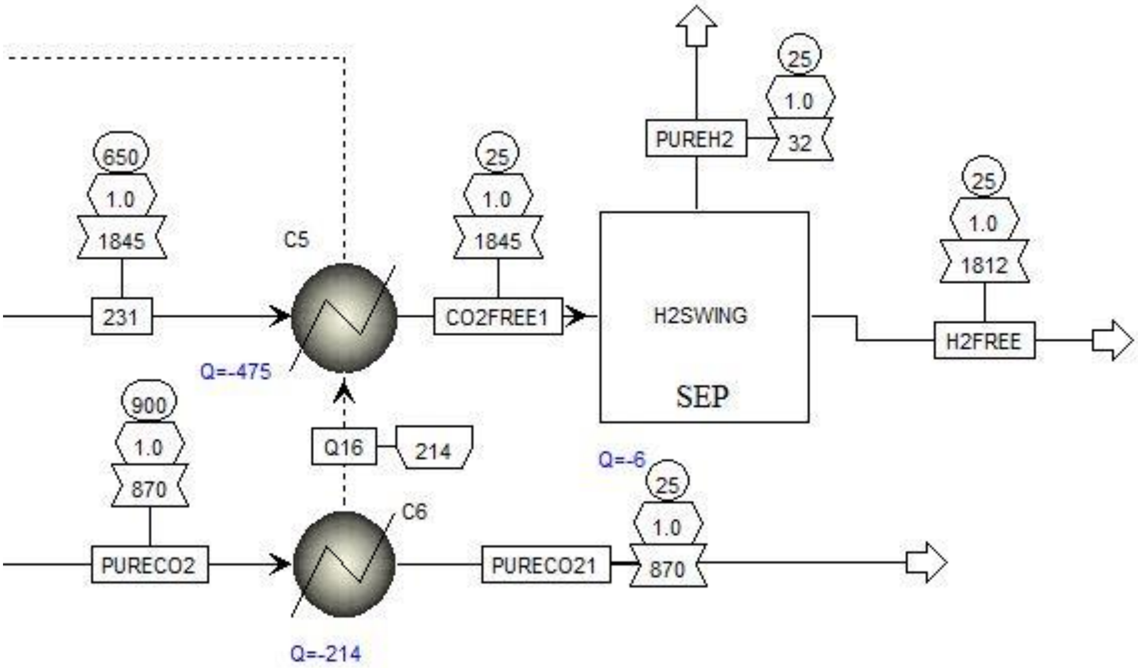
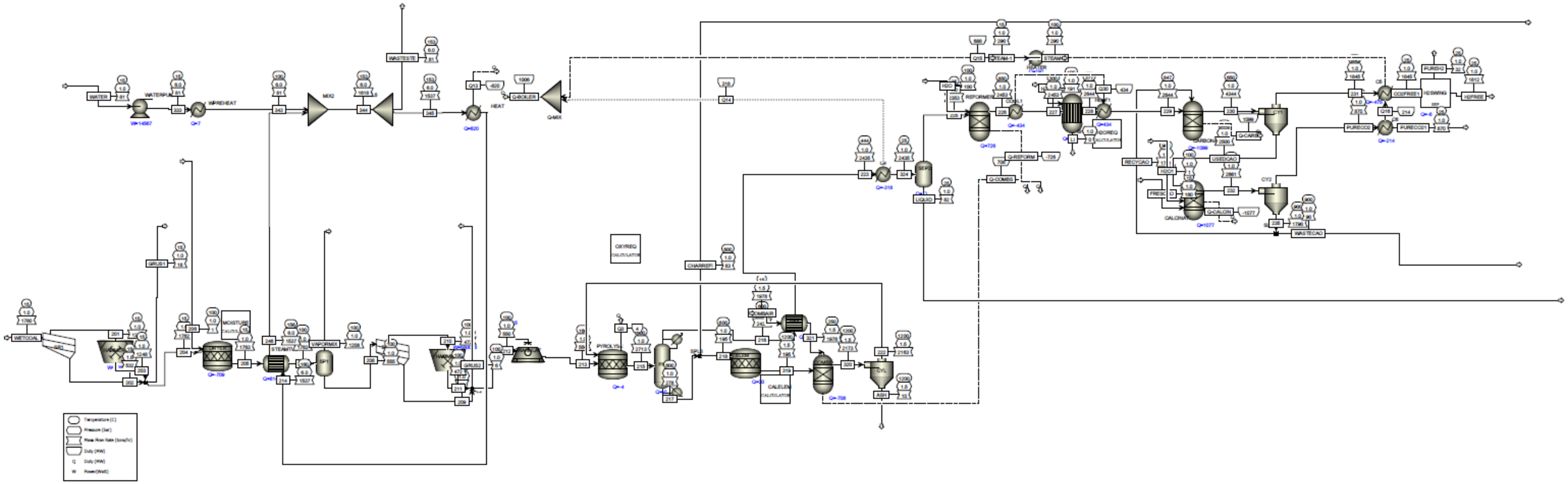


Fig 39: Heat Recovery and H2 Separation unit

Table 4 present the stream results

		WET COAL	STEAM-DRY	DRY-COAL	EXCESS-CHAR	LIQUID	CAO	PURECO2	H2
Mass Flow	TONS/HR	1780	81	549.5423	83.45252	81.70307	160	870.3749	32.35837
Mass Enthalpy	MW	-4752.92	-324.654	-551.546	45.77193	-106.152	-486.707	-1742.57	1.51E-14
Mass flow rate									
H2	TONS/HR	0	0	0	0	1.20E-05	0	0.00045	32.35837
CO	TONS/HR	0	0	0	0	0.00458	0	2.793017	0
CO2	TONS/HR	0	0	0	0	0.168121	0	864.4513	0
SO2	TONS/HR	0	0	0	0	0.006685	0	2.134137	0
SO3	TONS/HR	0	0	0	0	2.55E-08	0	1.51E-06	0
NO2	TONS/HR	0	0	0	0	0	0	0	0
NO	TONS/HR	0	0	0	0	3.36E-10	0	0	0
S	TONS/HR	0	0	0	0	0	0	0	0
H2O	TONS/HR	0	81	0	0	20.37918	0	0.995952	0
C	TONS/HR	0	0	0	0	0	0	0	0
C2H6	TONS/HR	0	0	0	0	0.000626	0	1.08E-31	0
CH4	TONS/HR	0	0	0	0	0.000942	0	5.54E-16	0
H2S	TONS/HR	0	0	0	0	0.000599	0	5.74E-05	0
C3H8	TONS/HR	0	0	0	0	0.000127	0	5.50E-47	0
C3H6-2	TONS/HR	0	0	0	0	9.86E-05	0	1.37E-40	0
O2	TONS/HR	0	0	0	0	7.89E-17	0	2.03E-09	0
N2	TONS/HR	0	0	0	0	0.033303	0	0	0
NH3	TONS/HR	0	0	0	0	0.00046	0	0	0
C10H8	TONS/HR	0	0	0	0	8.160822	0	1.31E-99	0
C6H6O	TONS/HR	0	0	0	0	22.69229	0	8.30E-63	0

TERT--01	TONS/HR	0	0	0	0	3.726848	0	2.75E-86	0
C2H4	TONS/HR	0	0	0	0	0.000966	0	4.24E-26	0
P-CRE-01	TONS/HR	0	0	0	0	26.52741	0	3.69E-78	0
CAO	TONS/HR	0	0	0	0	0	0	0	0
CACO3	TONS/HR	0	0	0	0	0	160	0	0
CASO3	TONS/HR	0	0	0	0	0	0	0	0
CASO4	TONS/HR	0	0	0	0	0	0	0	0
CAS	TONS/HR	0	0	0	0	0	0	0	0
COAL	TONS/HR	1780	0	549.5423	0	0	0	0	0
CHAR	TONS/HR	0	0	0	83.45252	0	0	0	0
ASH	TONS/HR	0	0	0	0	0	0	0	0


WATER
WATERPUMP
W=14567
WPREHEAT
Q=7
MIX2
WASTESTE
HEAT
Q=820
Q13
-820
Q-BOILER
1006
Q-MIX
GRUS1
WETCOAL
GR1
DRYER
Q=-709
MOISTURE
CALCULATOR
STEAMTUBE
VAPORMIX
SP1
GRUS2
PYROLYSIS
Q=-4
OXYREQ
CALCULATOR
CHARREFI
Q14
318
SPLIT
ELEM
COMBAIR
CALELEM
CALCULATOR
COMB
Q=-708
CYL
ASH
SEP2
LIQUID
Q=-318
H2O
REFORMER
Q=726
COOL1
Q=-434
Q-REFORM
-726
Q-COMBS
Q15
688
STEAM-1
HEATER
STEAM
Q30
434
HEAT1
H2OREQ
CALCULATOR
CARBONA
Q=-1099
Q-CARB
1099
USEDCAO
RECYCAO
H2O1
FRESC
CALCINAT
Q-CALCIN
-1077
Q=1077
CY1
CY2
WASTECAO
CO2FREE1
H2SWING
PUREH2
H2FREE
Q16
214
PURECO2
PURECO21
Q=-214
Temperature (C)
Pressure (bar)
Mass Flow Rate (tons/hr)
Duty (MW)
Q Duty (MW)
W Power(Watt)

3. Detailed simulation results of Option-3

The purpose of the present study to do a mass-energy balance for *H_2* production plant using Yallourn coal through pyrolysis. The production target of H_2 is 770 tons/day.

The entire plant has several sections from feed coal preparation to pyrolysis as well as downstream gas cleaning followed by pressure-swing adsorption of H_2 separation.

Several assumptions were made during simulation. The simulation begins with the mass balance from dry coal to pyrolysis product. Table 5 present the detailed mass balance of pyrolysis product and Balance between proximate and ultimate analysis of the parent coal.

Table 5: Balance of pyrolysis

SYMBOL	NAME	W%DAF	MOL%	C	H	N	S	O	ASH
DRY-CHAR	DRY-CHAR	50.6194	0.5062	46.3600	0.7277	0.4200	0.2400	0.2942	2.5800
H2O	WATER	5.5309	0.0031	0.0000	0.6189	0.0000	0.0000	4.9120	0.0000
C6H12O2	2-P,4-H,M	2.4253	0.0002	1.5047	0.2525	0.0000	0.0000	0.6681	0.0000
C6H6O	PHENOL	4.2426	0.0005	3.2488	0.2726	0.0000	0.0000	0.7213	0.0000
C7H8O	P-CRESOL	4.8614	0.0004	3.7797	0.3172	0.0000	0.0000	0.7646	0.0000
C10H8	NAPTHENE	1.5127	0.0001	1.4176	0.0952	0.0000	0.0000	0.0000	0.0000
H2	HYDROGEN	1.4200	0.0070	0.0000	1.4200	0.0000	0.0000	0.0000	0.0000
CH4	METHANE	3.0000	0.0019	2.2461	0.7539	0.0000	0.0000	0.0000	0.0000
C2H4	ETHYLENE	0.5830	0.0002	0.4992	0.0838	0.0000	0.0000	0.0000	0.0000
C2H6	ETAHNE	0.2271	0.0001	0.1814	0.0457	0.0000	0.0000	0.0000	0.0000
C3H6	PROPYLENE	0.0100	0.0000	0.0086	0.0014	0.0000	0.0000	0.0000	0.0000
C3H8	PROPANE	0.0100	0.0000	0.0082	0.0018	0.0000	0.0000	0.0000	0.0000
CO	CO	11.5984	0.0041	4.9735	0.0000	0.0000	0.0000	6.6250	0.0000
CO2	CO2	13.8691	0.0032	3.7851	0.0000	0.0000	0.0000	10.0840	0.0000
NH3	NH3	0.0399	0.0000	0.0000	0.0071	0.0328	0.0000	0.0000	0.0000
H2S	H2S	0.0501	0.0000	0.0000	0.0030	0.0000	0.0471	0.0000	0.0000
	TOTAL	100.0000	0.5270	68.0126	4.6008	0.4528	0.2871	24.0691	2.5800

PROXIMATE	COAL	Ultimate	coal
ASH	2.58	C	68.01

MOISTURE	5.53	H	4.60
FC	48.04	N	0.45
VOL	43.85	S	0.29
		O	24.07
		ASH	2.58

The parent coal was considered with 60% moisture content. The moisture content reduced to 12 after steam tube dryer. The moisture is further reduced to 5.53 after briquette process.

The details of the individual unit operation are as follows:

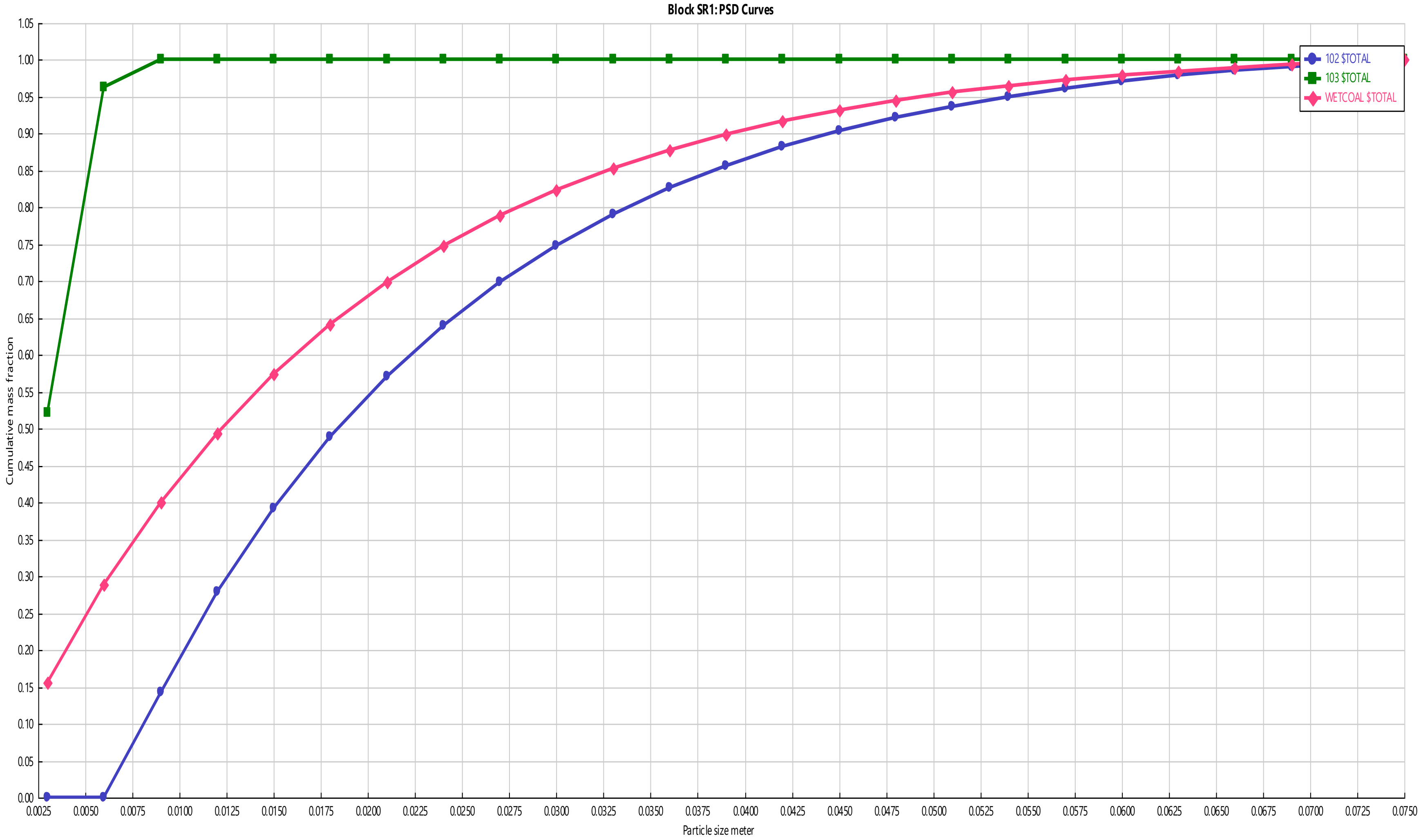
Coal storage

Coal storage is not a process operation and not modelled.

Primary screening and Hammering

The raw coal is considered with 60% moisture with a PSD as described before. This wet coal goes to a primary hammer mill followed by a preliminary screening.

The primary screener is modelled as default SCREEN with a cut size of 0.0063 m (Coarse-split) to meet the primary Hammer mill requirement. The separation stage and efficiency is presented in **Fig 41**.


Block SR1: PSD Curves
Cumulative mass fraction
1.05
1.00
0.95
0.90
0.85
0.80
0.75
0.70
0.65
0.60
0.55
0.50
0.45
0.40
0.35
0.30
0.25
0.20
0.15
0.10
0.05
0.00
0.0025
0.0050
0.0075
0.0100
0.0125
0.0150
0.0175
0.0200
0.0225
0.0250
0.0275
0.0300
0.0325
0.0350
0.0375
0.0400
0.0425
0.0450
0.0475
0.0500
0.0525
0.0550
0.0575
0.0600
0.0625
0.0650
0.0675
0.0700
0.0725
0.0750
Particle size meter
102 $TOTAL
103 $TOTAL
WETCOAL $TOTAL

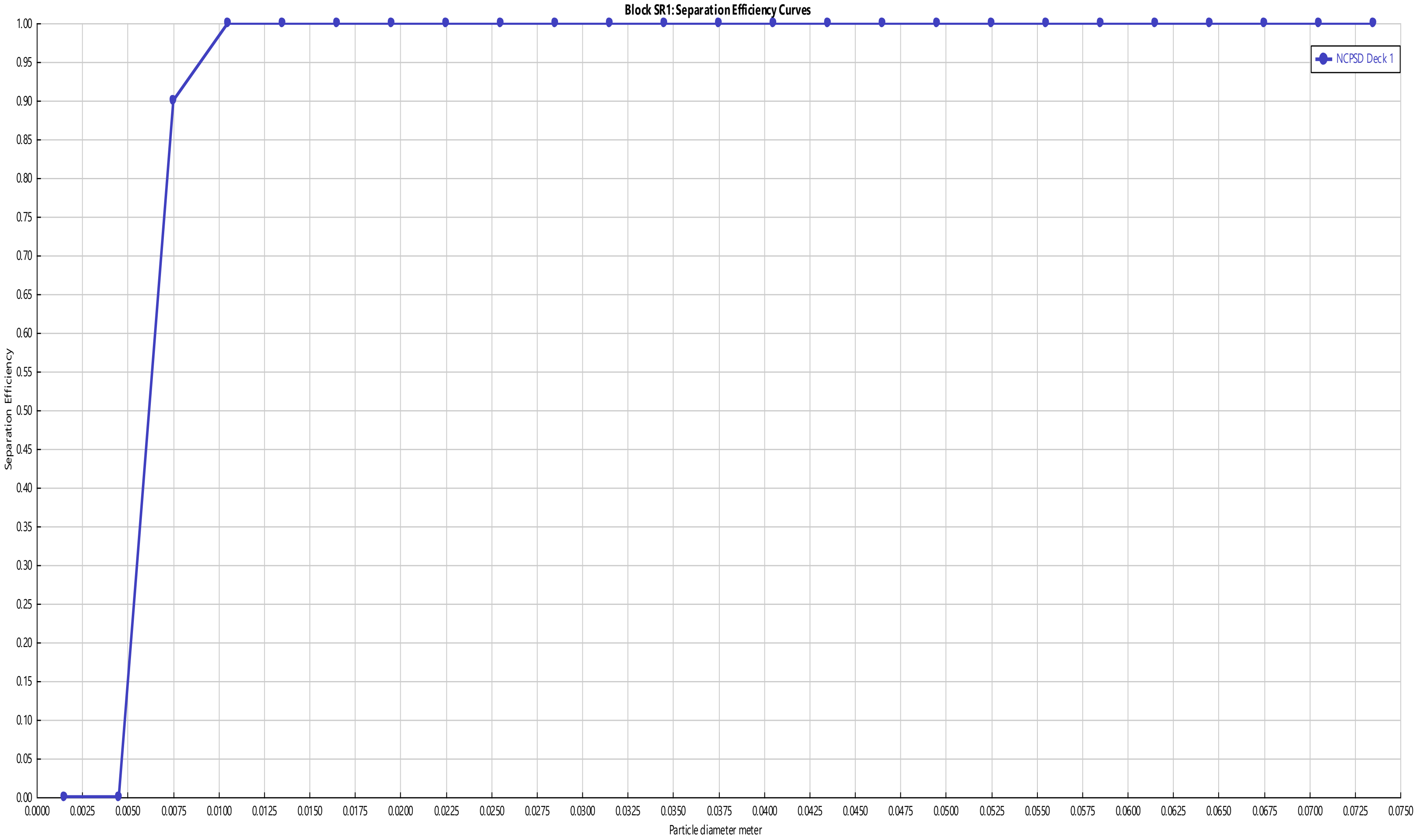
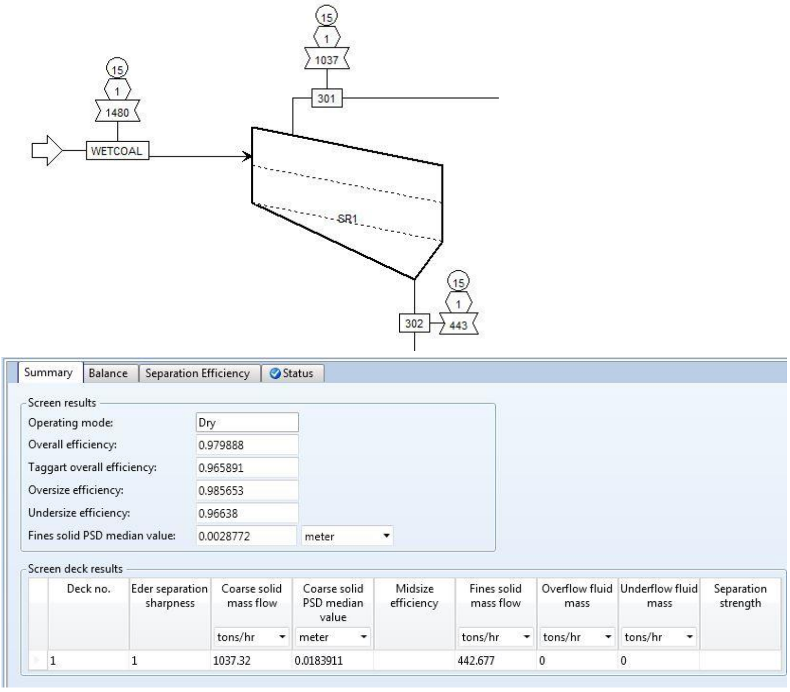


Fig 41: Primary screening separation and efficiency

The details of the primary SCREENING machine is presented in **Fig 42.**



Screen results

Operating mode: Dry
Overall efficiency: 0.979888
Taggart overall efficiency: 0.965891
Oversize efficiency: 0.985653
Undersize efficiency: 0.96638
Fines solid PSD median value: 0.0028772 meter

Screen deck results

Deck no.	Eder separation sharpness	Coarse solid mass flow tons/hr	Coarse solid PSD median value meter	Midsize efficiency	Fines solid mass flow tons/hr	Overflow fluid mass tons/hr	Underflow fluid mass tons/hr	Separation strength
1	1	1037.32	0.0183911		442.677	0	0	

Fig 42: Details of primary screener

The hammer mill is considered as a HAMMER CRUSHER with default inbuilt model. The PSD curve shape, i.e. RRSB distribution function using dispersion parameter=2, D63 particle size adjusted to suit ~98 wt% particles <9.3 mm and secondary split. The open cut method with Bands law (HG-index 130) is used.

The PSD after primary hammering is presented in **Fig 43.**

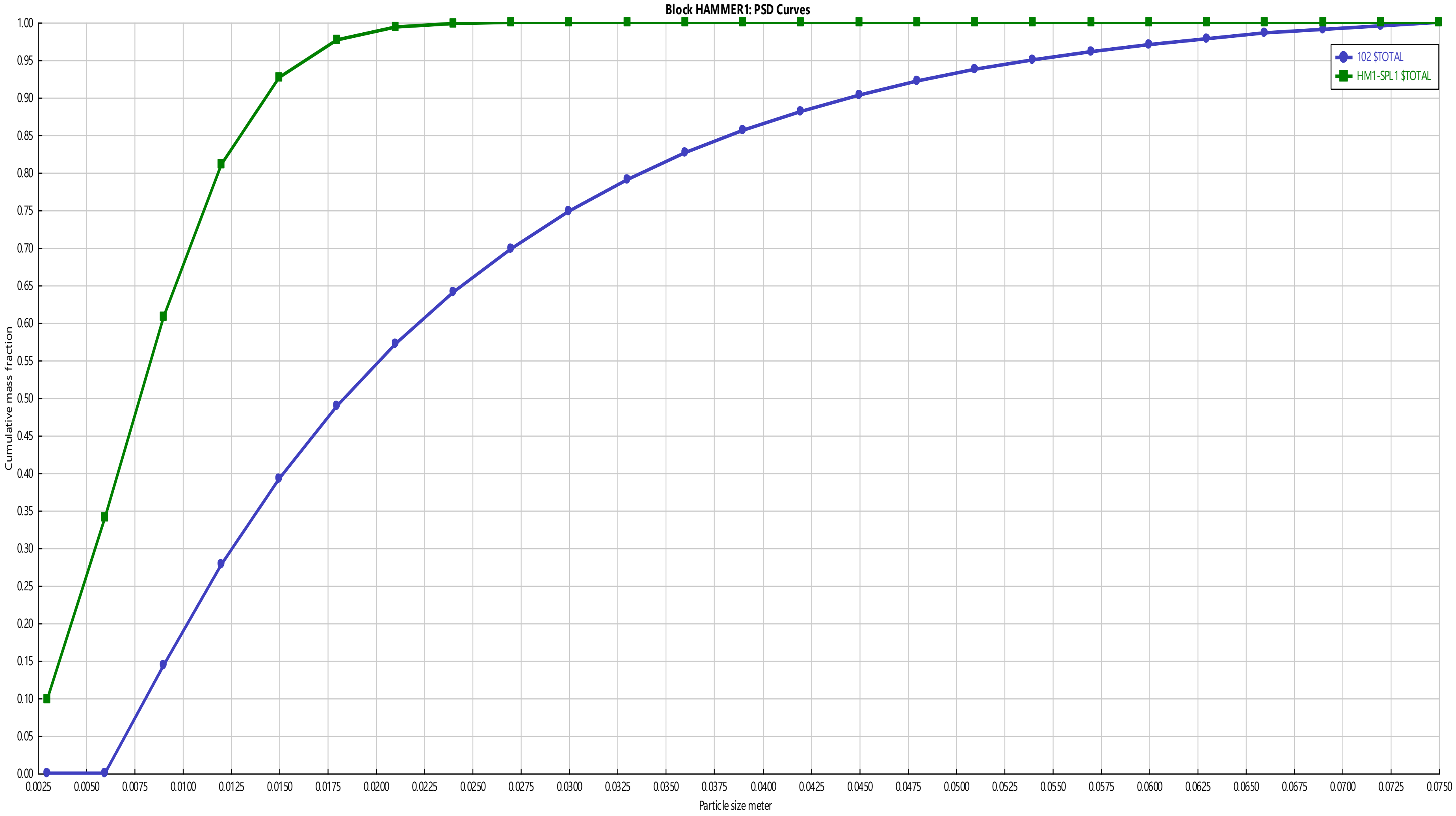
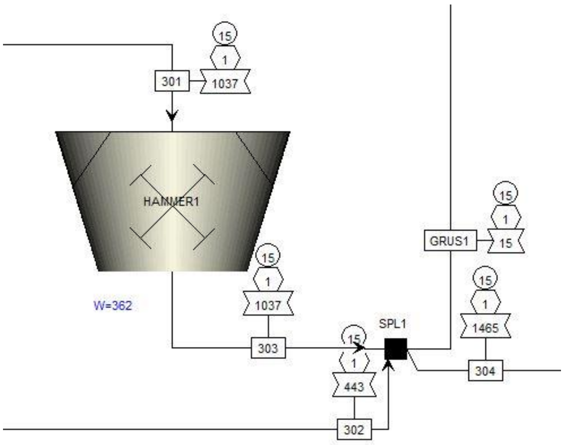
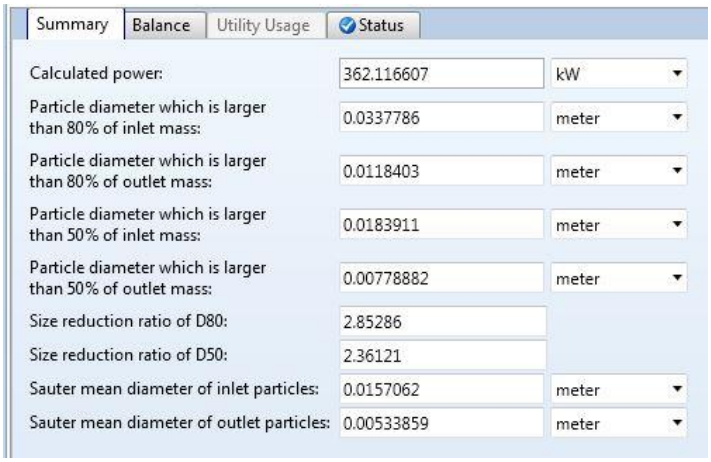


Fig 43: PSD after primary hammering

The details of the primary Hammer machine is presented in **Fig 44.**





Calculated power:	362.116607	kW
Particle diameter which is larger than 80% of inlet mass:	0.0337786	meter
Particle diameter which is larger than 80% of outlet mass:	0.0118403	meter
Particle diameter which is larger than 50% of inlet mass:	0.0183911	meter
Particle diameter which is larger than 50% of outlet mass:	0.00778882	meter
Size reduction ratio of D80:	2.85286	
Size reduction ratio of D50:	2.36121	
Sauter mean diameter of inlet particles:	0.0157062	meter
Sauter mean diameter of outlet particles:	0.00533859	meter

Fig 44: Details of Primary Hammer

**1% of raw coal was considered as grus and was not accounted in the mainstream. An FSPLIT was used for this purpose.

Steam Tube dryer

The steam tube dryer is modelled with 3 different default inbuilt unit operation unit. An RStoic is used to convert the non-conventional moisture present in coal to convert conventional pure water at 15°C based on Eq (1):

WETCOAL--> 0.0555084 H_2O …….. (1)

The pure conventional water is passed through a heat exchanger (HX) with a specification of vapour fraction 1 at the outlet of the cold fluid. All the water that is vaporised in the HX unit is separate in a splitter (SP1). **1% of entrainment of coal was considered in the water vapour.** The overall unit of the dryer is presented in **Fig 45.**

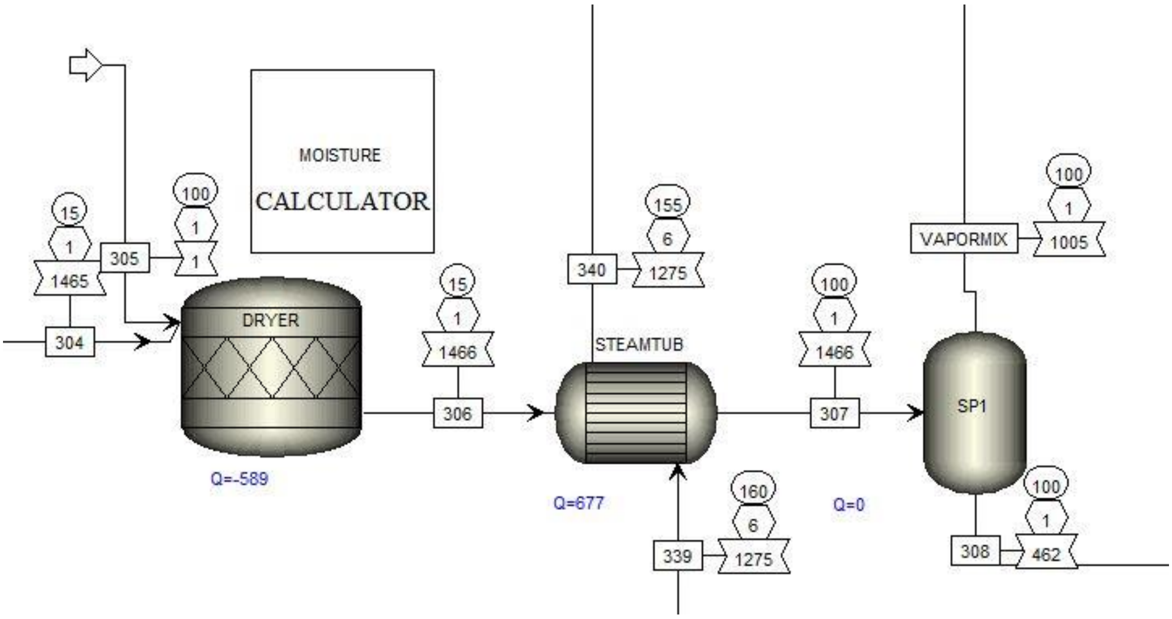
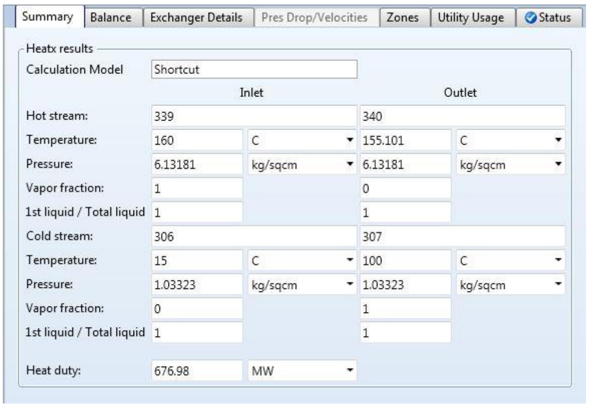


Fig 45: Steam tube dryer

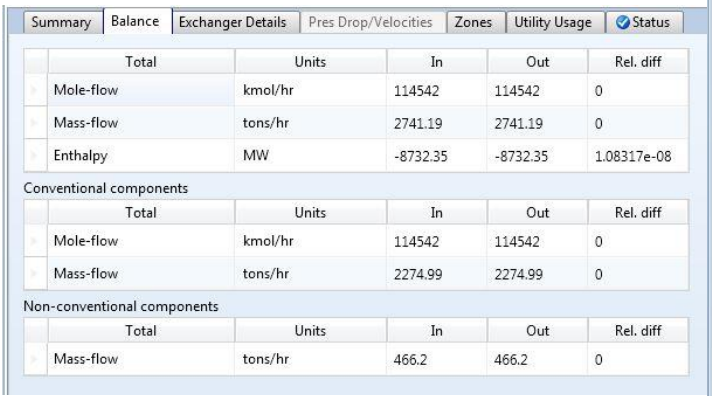
A stream of air 305 is introduced in the DRYER to pull out the moisture from the tubes. 160°C and 5 barg steam were used in the hot stream of the HX. The steam requirement was adjusted such a way that the temperature drop after HX of the hot stream is 155°C. The details of the HX unit is presented in **Fig 46.**



Heatx results

Calculation Model: Shortcut

	Inlet		Outlet	
Hot stream:	339		340	
Temperature:	160	C	155.101	C
Pressure:	6.13181	kg/sqcm	6.13181	kg/sqcm
Vapor fraction:	1		0	
1st liquid / Total liquid	1		1	
Cold stream:	306		307	
Temperature:	15	C	100	C
Pressure:	1.03323	kg/sqcm	1.03323	kg/sqcm
Vapor fraction:	0		1	
1st liquid / Total liquid	1		1	
Heat duty:	676.98	MW		



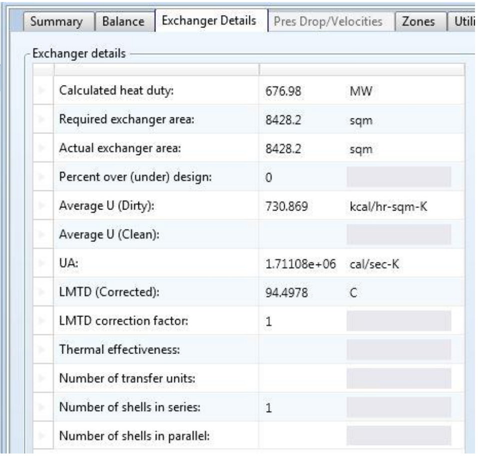
Total	Units	In	Out	Rel. diff
Mole-flow	kmol/hr	114542	114542	0
Mass-flow	tons/hr	2741.19	2741.19	0
Enthalpy	MW	-8732.35	-8732.35	1.08317e-08

Conventional components

Total	Units	In	Out	Rel. diff
Mole-flow	kmol/hr	114542	114542	0
Mass-flow	tons/hr	2274.99	2274.99	0

Non-conventional components

Total	Units	In	Out	Rel. diff
Mass-flow	tons/hr	466.2	466.2	0



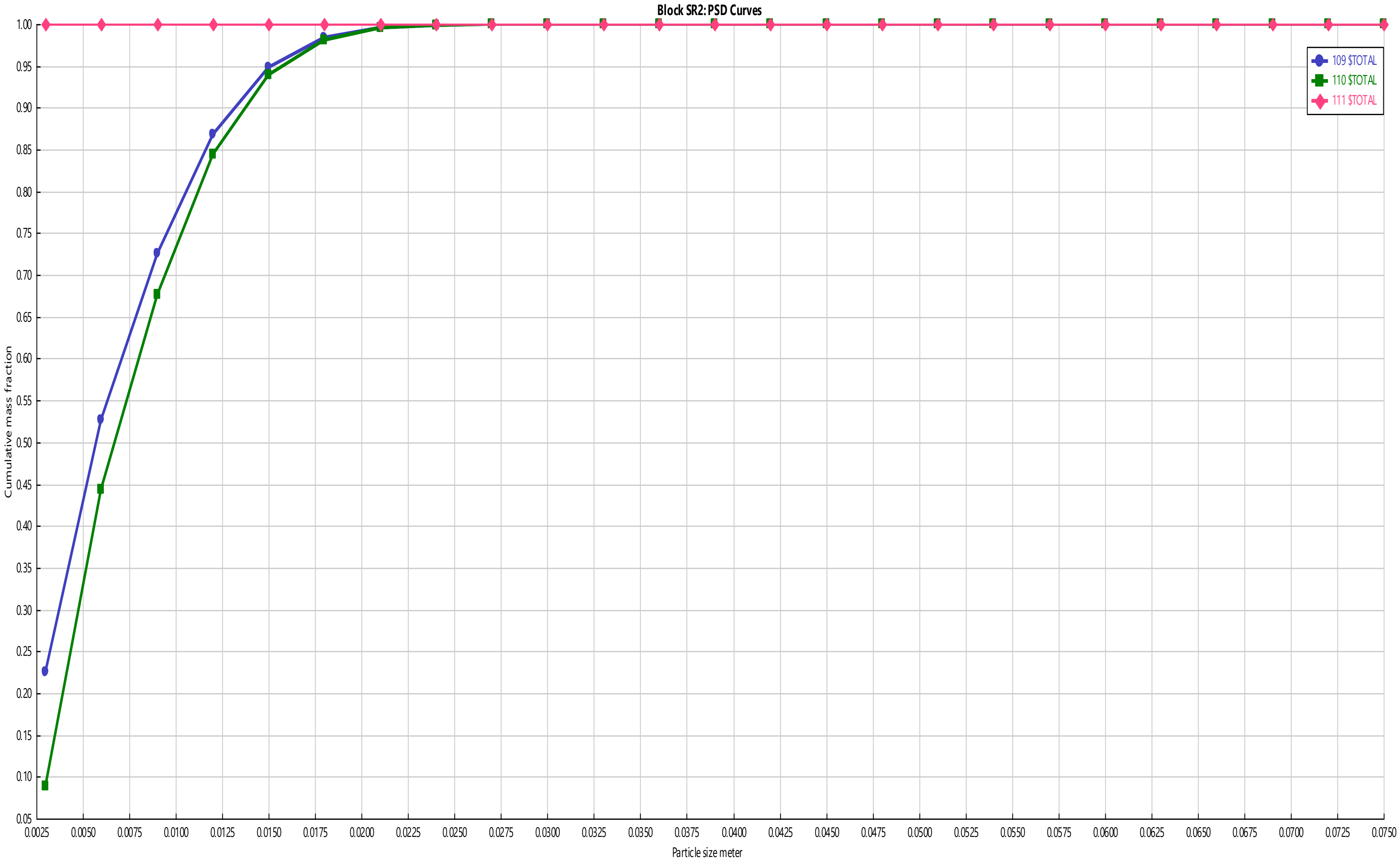
Exchanger details

Calculated heat duty:	676.98	MW
Required exchanger area:	8428.2	sqm
Actual exchanger area:	8428.2	sqm
Percent over (under) design:	0	
Average U (Dirty):	730.869	kcal/hr-sqm-K
Average U (Clean):		
UA:	1.71108e+06	cal/sec-K
LMTD (Corrected):	94.4978	C
LMTD correction factor:	1	
Thermal effectiveness:		
Number of transfer units:		
Number of shells in series:	1	
Number of shells in parallel:		

Fig 46: Details of Steam tube dryer unit

Secondary screening and Hammering

The dry coal goes to a secondary screening and hammering for further refinement. The secondary screener is modelled as default SCREEN with a cut size of 0.002 m (Coarse-split) to meet the secondary Hammer mill requirement. The separation stage and efficiency is presented in **Fig 47**.


Block SR2: PSD Curves
1.00
0.95
0.90
0.85
0.80
0.75
0.70
0.65
0.60
0.55
0.50
0.45
0.40
0.35
0.30
0.25
0.20
0.15
0.10
0.05
Cumulative mass fraction
0.0025
0.0050
0.0075
0.0100
0.0125
0.0150
0.0175
0.0200
0.0225
0.0250
0.0275
0.0300
0.0325
0.0350
0.0375
0.0400
0.0425
0.0450
0.0475
0.0500
0.0525
0.0550
0.0575
0.0600
0.0625
0.0650
0.0675
0.0700
0.0725
0.0750
Particle size meter
109 $TOTAL
110 $TOTAL
111 $TOTAL

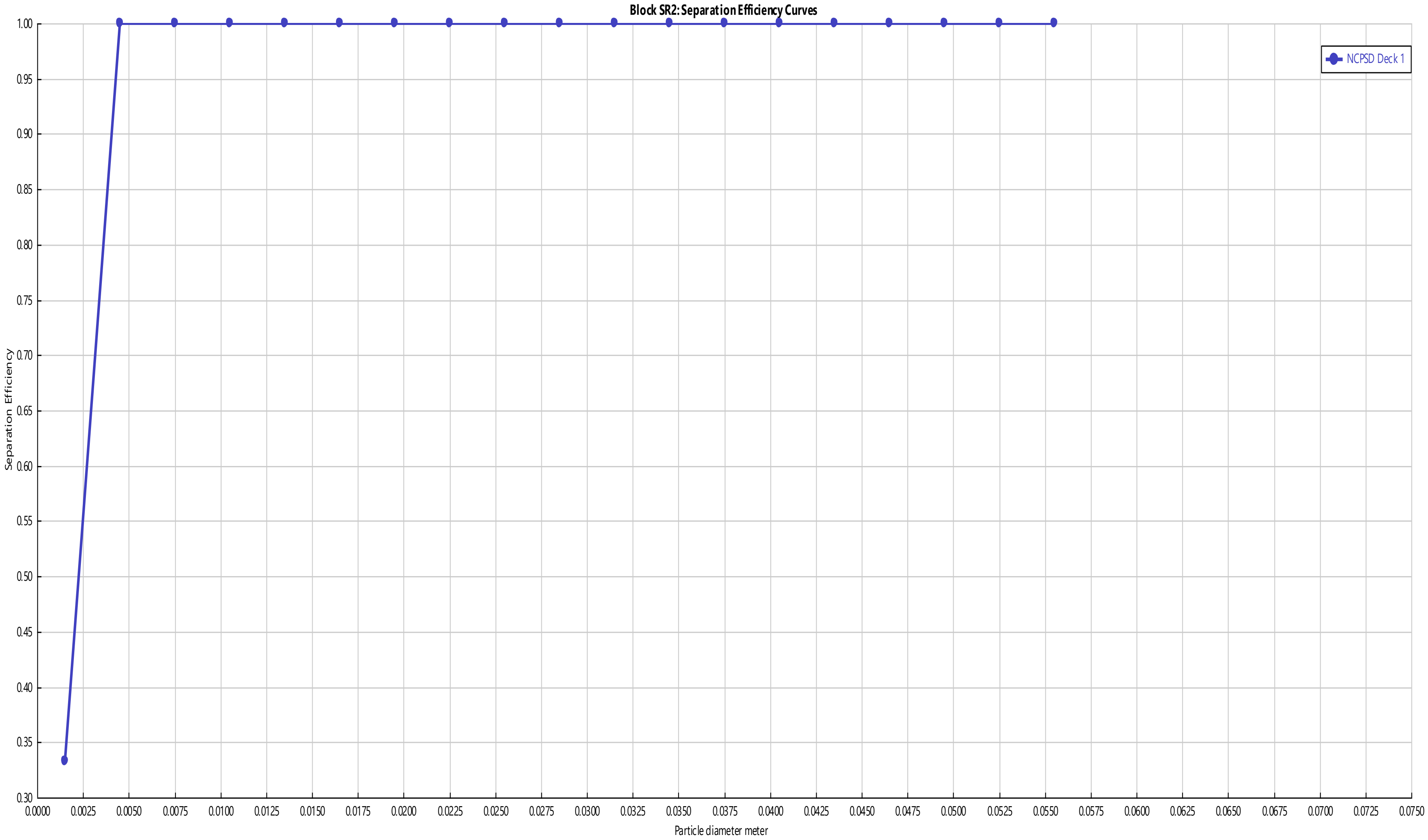
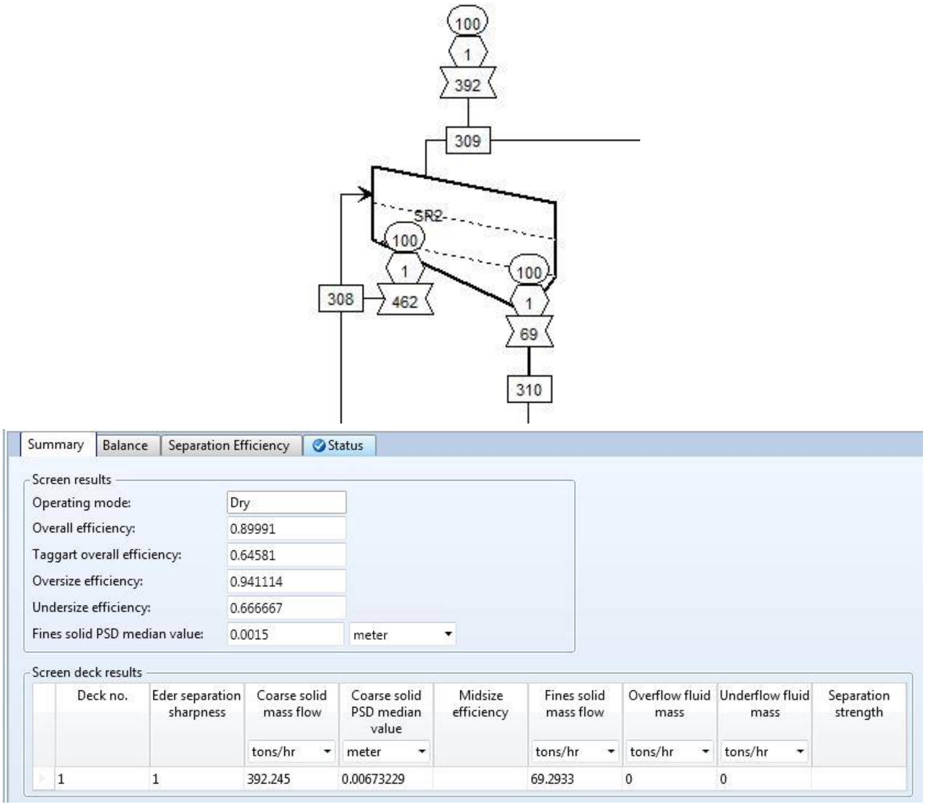

Block SR2: Separation Efficiency Curves
NCPSD Deck 1
Separation Efficiency
1.00
0.95
0.90
0.85
0.80
0.75
0.70
0.65
0.60
0.55
0.50
0.45
0.40
0.35
0.30
Particle diameter meter
0.0000
0.0025
0.0050
0.0075
0.0100
0.0125
0.0150
0.0175
0.0200
0.0225
0.0250
0.0275
0.0300
0.0325
0.0350
0.0375
0.0400
0.0425
0.0450
0.0475
0.0500
0.0525
0.0550
0.0575
0.0600
0.0625
0.0650
0.0675
0.0700
0.0725
0.0750

Fig 47: Secondary screening separation and efficiency

The details of the primary SCREENING machine is presented in **Fig 48.**



Screen results

Operating mode: Dry

Overall efficiency: 0.89991

Taggart overall efficiency: 0.64581

Oversize efficiency: 0.941114

Undersize efficiency: 0.666667

Fines solid PSD median value: 0.0015 meter

Screen deck results

Deck no.	Eder separation sharpness	Coarse solid mass flow	Coarse solid PSD median value	Midsize efficiency	Fines solid mass flow	Overflow fluid mass	Underflow fluid mass	Separation strength
		tons/hr	meter		tons/hr	tons/hr	tons/hr	
1	1	392.245	0.00673229		69.2933	0	0	

Fig 48: Details of secondary screener

The secondary hammer mill is considered as a HAMMER CRUSHER with default inbuilt model. The PSD curve shape i.e. RRSB distribution function using dispersion parameter=2, D63 particle size adjusted to suit ~98 wt% particles<2 mm and secondary split.

The PSD after primary hammering is presented in **Fig 49.**

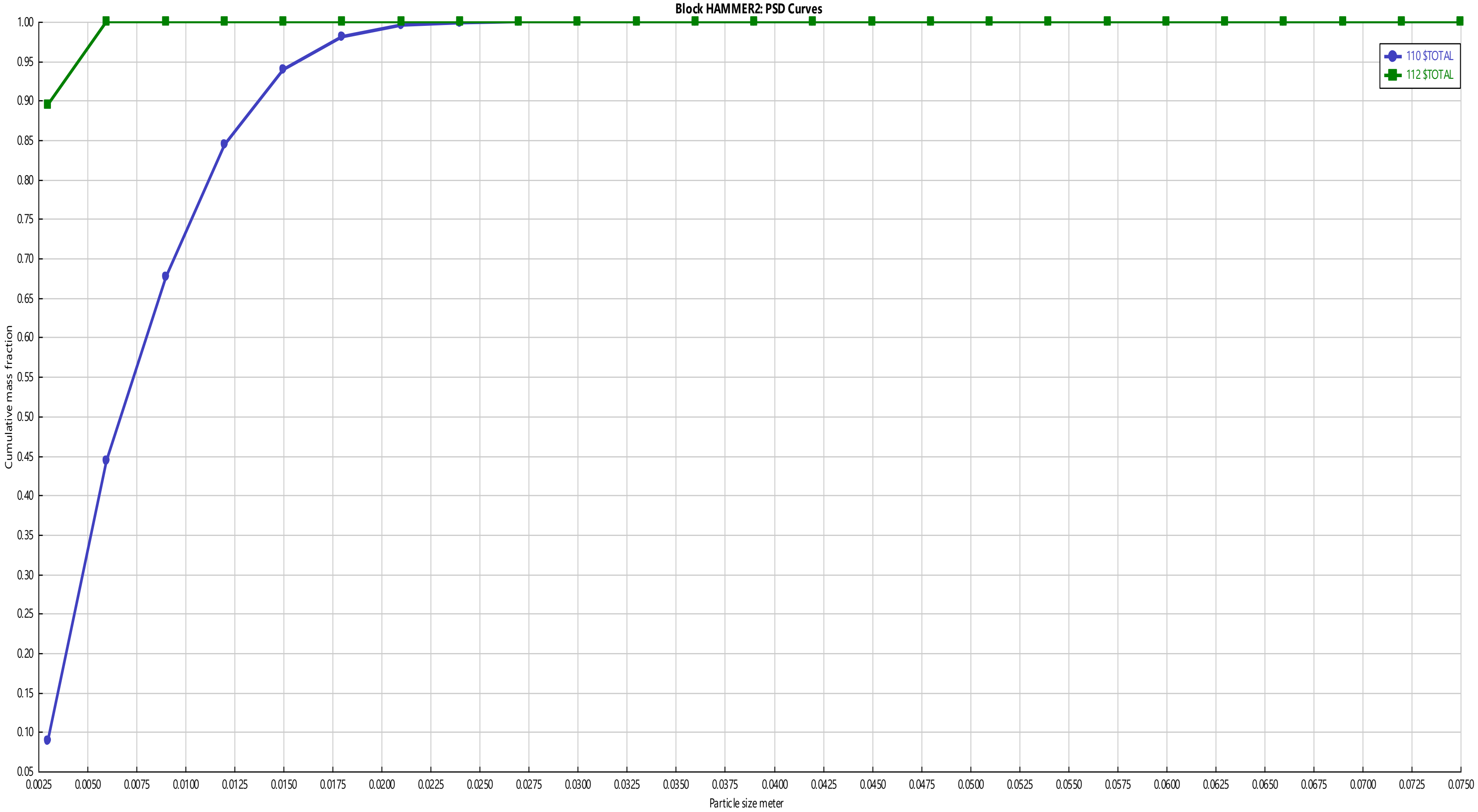
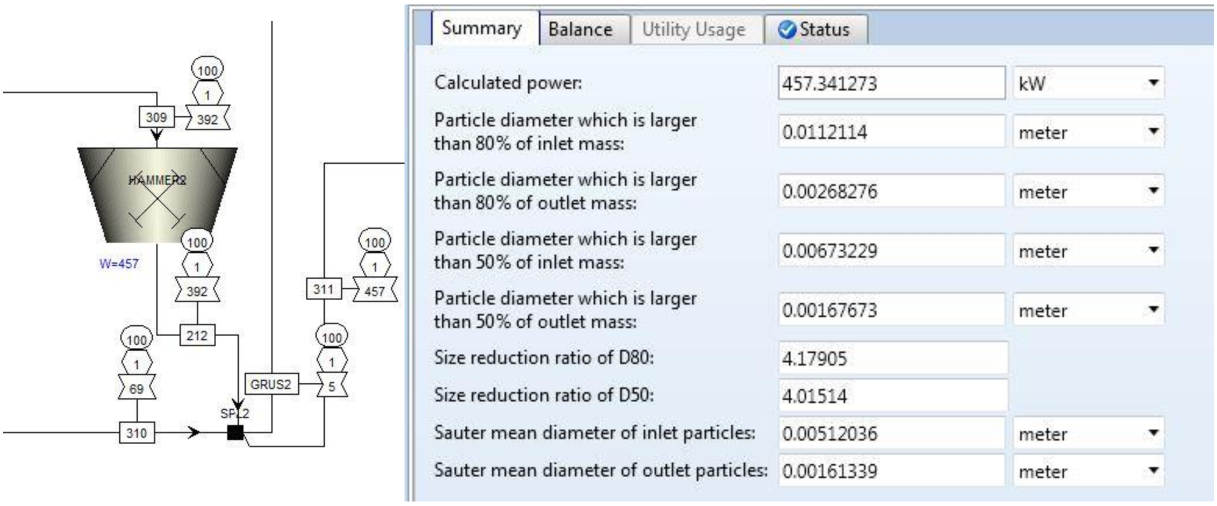


Fig 49: PSD of Secondary Hammer

The details of the secondary Hammer machine is presented in **Fig 50.**



Calculated power:	457.341273	kW
Particle diameter which is larger than 80% of inlet mass:	0.0112114	meter
Particle diameter which is larger than 80% of outlet mass:	0.00268276	meter
Particle diameter which is larger than 50% of inlet mass:	0.00673229	meter
Particle diameter which is larger than 50% of outlet mass:	0.00167673	meter
Size reduction ratio of D80:	4.17905	
Size reduction ratio of D50:	4.01514	
Sauter mean diameter of inlet particles:	0.00512036	meter
Sauter mean diameter of outlet particles:	0.00161339	meter

Fig 50: Details of Secondary Hammer

**1% of raw coal was considered as grus and was not accounted in the mainstream. An FSPLIT was used for this purpose.

Briquette unit

The briquette unit is modelled as default granulator model of process simulation package. The briquette is considered as 62 mm sphere diameter for an ellipsoid containing a dimension of 79x80x30 equivalent to 126 cm^3. The PSD is shown in **Fig 51** and the details in **Fig 52.**

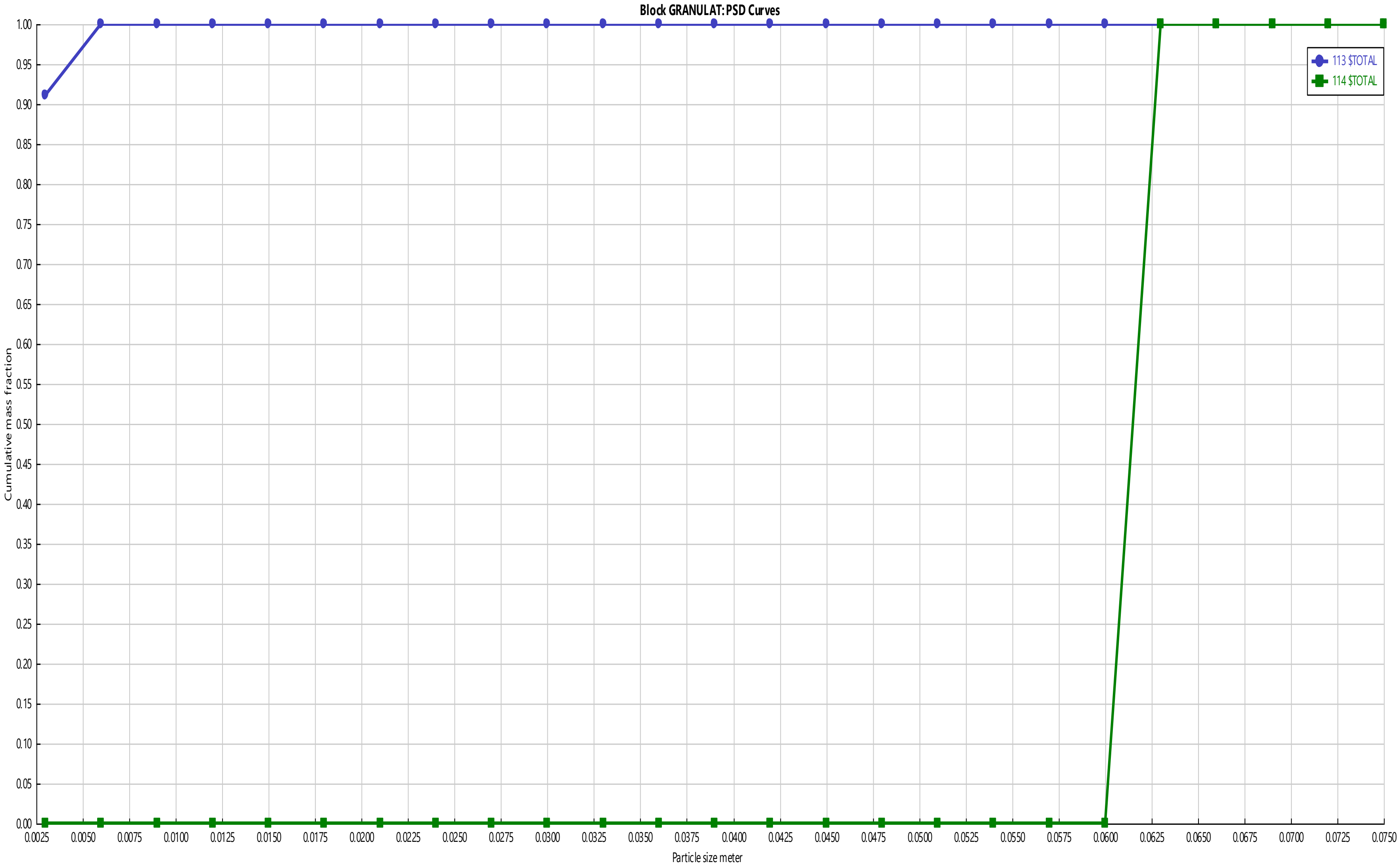
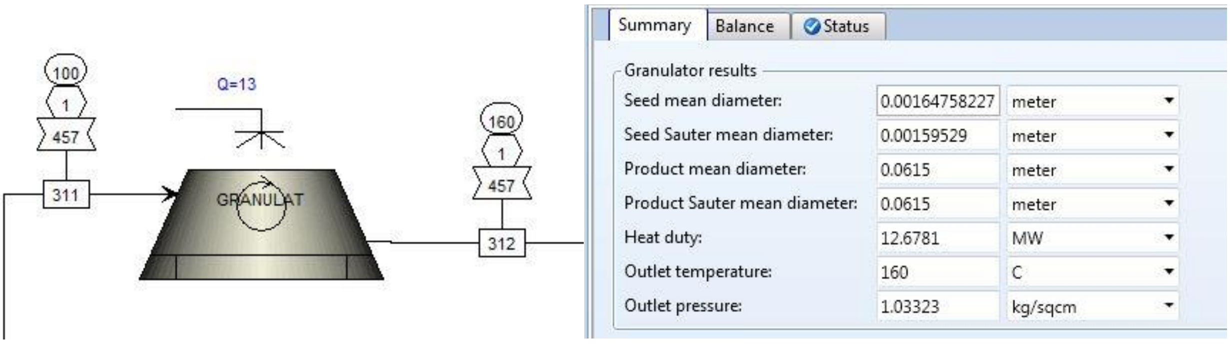


Fig 51: PSD in Granulator



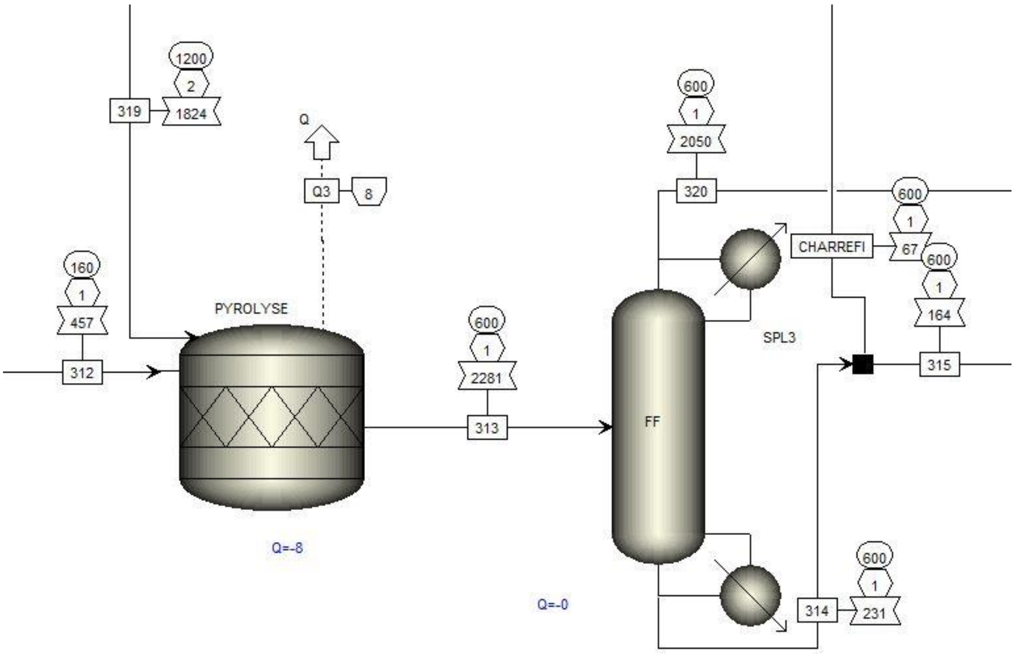
Granulator results

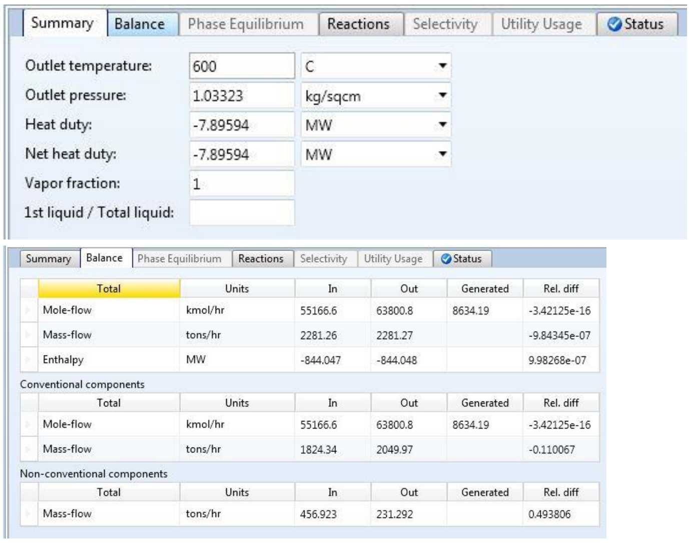
Seed mean diameter:	0.00164758227	meter
Seed Sauter mean diameter:	0.00159529	meter
Product mean diameter:	0.0615	meter
Product Sauter mean diameter:	0.0615	meter
Heat duty:	12.6781	MW
Outlet temperature:	160	C
Outlet pressure:	1.03323	kg/sqcm

Fig 52: Details of Granulator unit

Pyrolyser unit

The Pyrolyser unit is modelled as an RStoic reactor with a specific reactor temperature of 600°C and a component splitter. The heat requirement for the unit is satisfied by direct passing hot flue gas from the combustor. We assumed the flue gases from the combustor do not react with the pyrolysis products. The heat requirement is adjusted through the extent of char burning. It was found 71% of the char should be burnt to meet the energy requirement of the pyrolyzer. Excess heat energy was considered in the reactor. The details of the pyrolyzer are presented in **Fig 52.**





Outlet temperature: 600 C
Outlet pressure: 1.03323 kg/sqcm
Heat duty: -7.89594 MW
Net heat duty: -7.89594 MW
Vapor fraction: 1
1st liquid / Total liquid:

Total	Units	In	Out	Generated	Rel. diff
Mole-flow	kmol/hr	55166.6	63800.8	8634.19	-3.42125e-16
Mass-flow	tons/hr	2281.26	2281.27		-9.84345e-07
Enthalpy	MW	-844.047	-844.048		9.98268e-07

Conventional components

Total	Units	In	Out	Generated	Rel. diff
Mole-flow	kmol/hr	55166.6	63800.8	8634.19	-3.42125e-16
Mass-flow	tons/hr	1824.34	2049.97		-0.110067

Non-conventional components

Total	Units	In	Out	Generated	Rel. diff
Mass-flow	tons/hr	456.923	231.292		0.493806

Fig 53: Details of the pyrolyzer unit

Combustor unit

The combustor unit was modelled as an RStoic, RGibbs and a Cyclone separator presented in **Fig 54**. The RStoic is used to convert the char into its preliminary constitution. Standard combustion reactions were considered in the RGibbs reactor. The Ash was separated by Cyclone separator.

The combustor operates at 1200°C, 2 bar pressure. The air requirement for the char combustion was calculated with a calculator block. The air was preheated to 250°C through the hot pyrolysis gases by a heat exchanger (HX1) with a specific cold outlet temperature of 250°C. 71% of the char from the pyrolysis unit was combusted to meet the energy requirement of the pyrolyzer and other accessories.

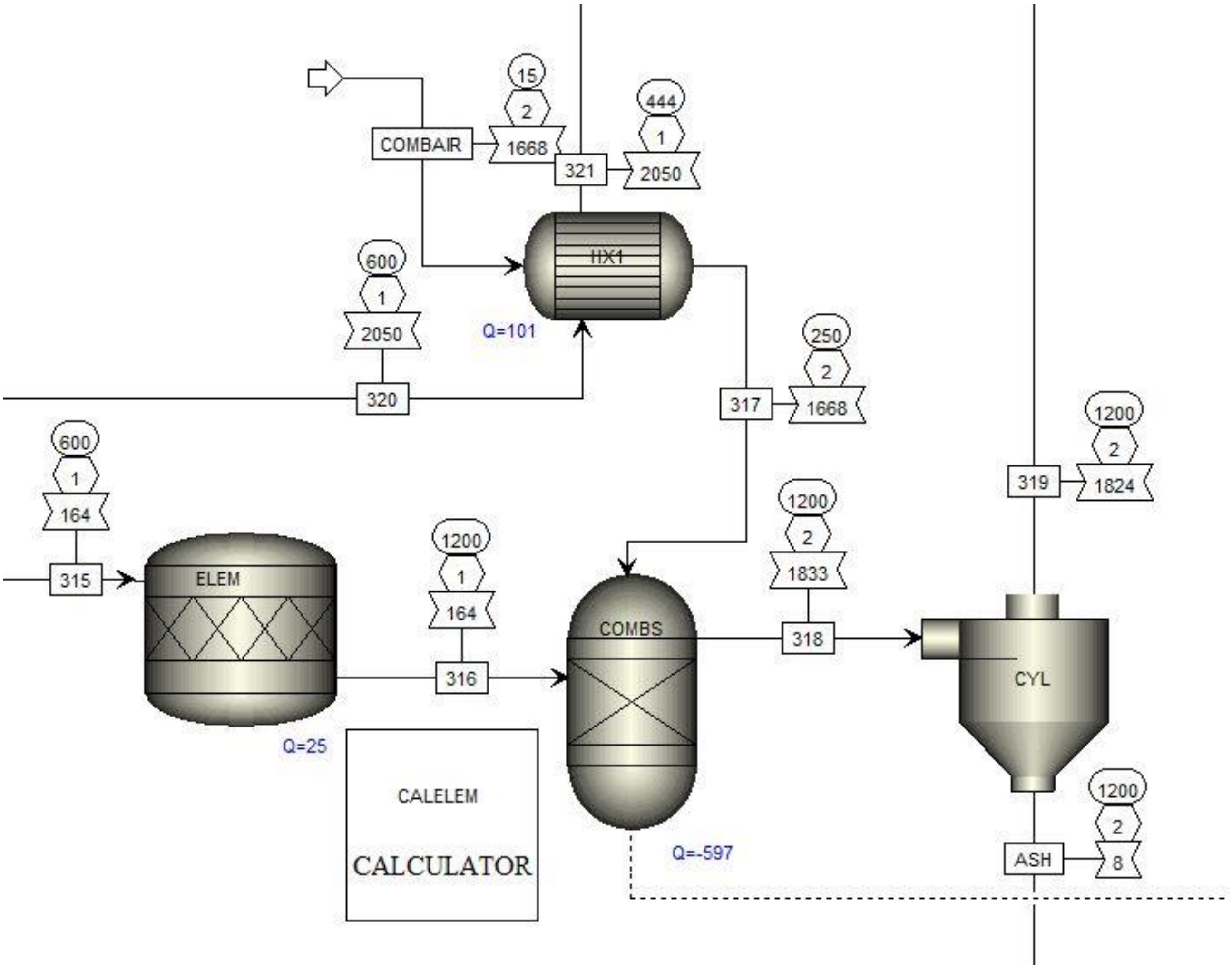
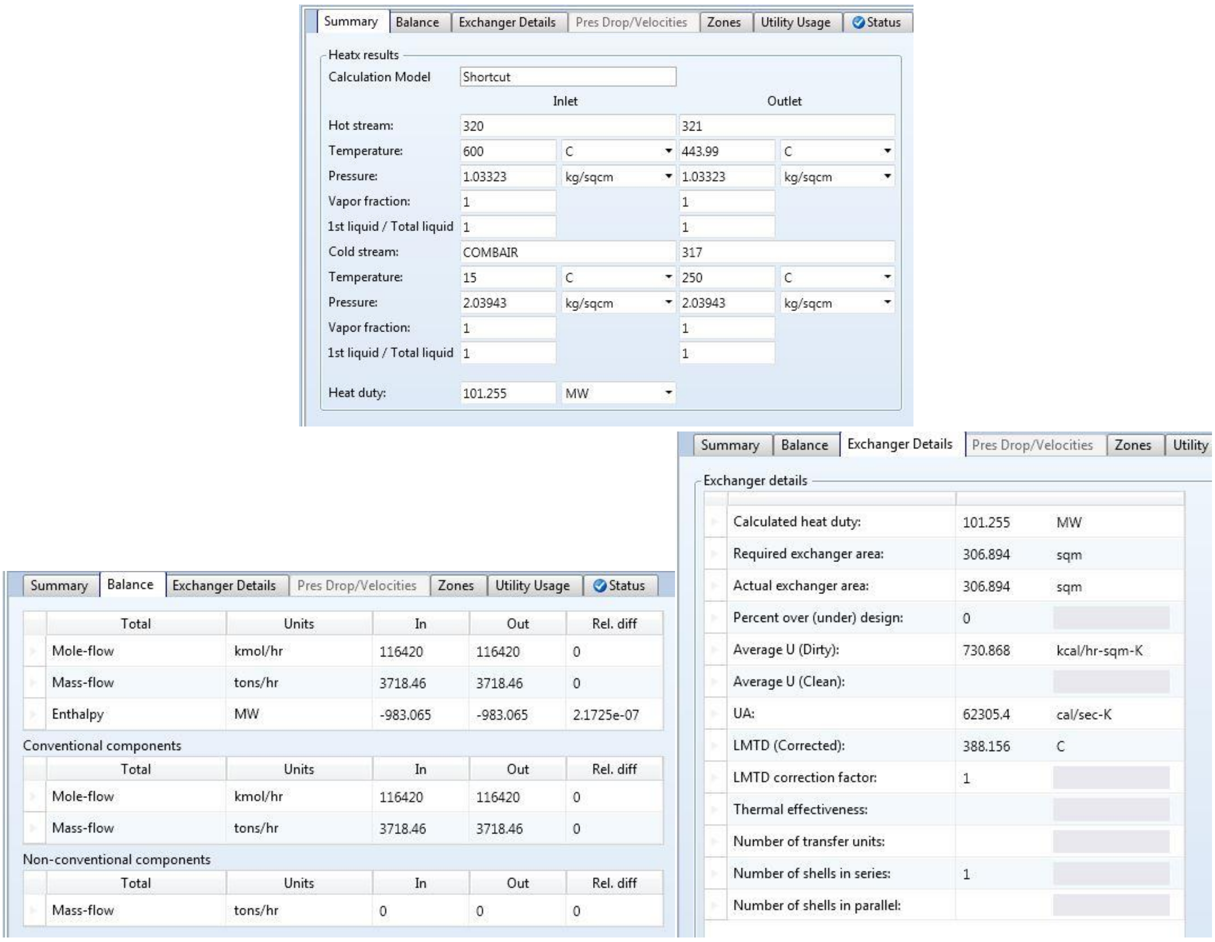


Fig 54: Combustor unit

The details of the HX1 unit is presented in **Fig 55.**



	Inlet		Outlet	
Calculation Model	Shortcut			
Hot stream:	320		321	
Temperature:	600	C	443.99	C
Pressure:	1.03323	kg/sqcm	1.03323	kg/sqcm
Vapor fraction:	1		1	
1st liquid / Total liquid	1		1	
Cold stream:	COMBAIR		317	
Temperature:	15	C	250	C
Pressure:	2.03943	kg/sqcm	2.03943	kg/sqcm
Vapor fraction:	1		1	
1st liquid / Total liquid	1		1	
Heat duty:	101.255	MW		

Total	Units	In	Out	Rel. diff
Mole-flow	kmol/hr	116420	116420	0
Mass-flow	tons/hr	3718.46	3718.46	0
Enthalpy	MW	-983.065	-983.065	2.1725e-07

Conventional components

Total	Units	In	Out	Rel. diff
Mole-flow	kmol/hr	116420	116420	0
Mass-flow	tons/hr	3718.46	3718.46	0

Non-conventional components

Total	Units	In	Out	Rel. diff
Mass-flow	tons/hr	0	0	0

Exchanger details

Calculated heat duty:	101.255	MW
Required exchanger area:	306.894	sqm
Actual exchanger area:	306.894	sqm
Percent over (under) design:	0	
Average U (Dirty):	730.868	kcal/hr-sqm-K
Average U (Clean):		
UA:	62305.4	cal/sec-K
LMTD (Corrected):	388.156	C
LMTD correction factor:	1	
Thermal effectiveness:		
Number of transfer units:		
Number of shells in series:	1	
Number of shells in parallel:		

Fig 55. Details of air preheater for char combustor

Pyrolysis Oil extraction unit

The stream of pyrolysis gases was further cooled down to 25°C with a simple Heater to calculate the energy availability to produce steam for the dryer unit. The liquid-vapour was separated with a FLASH unit. **Fig 56** presents the scheme.

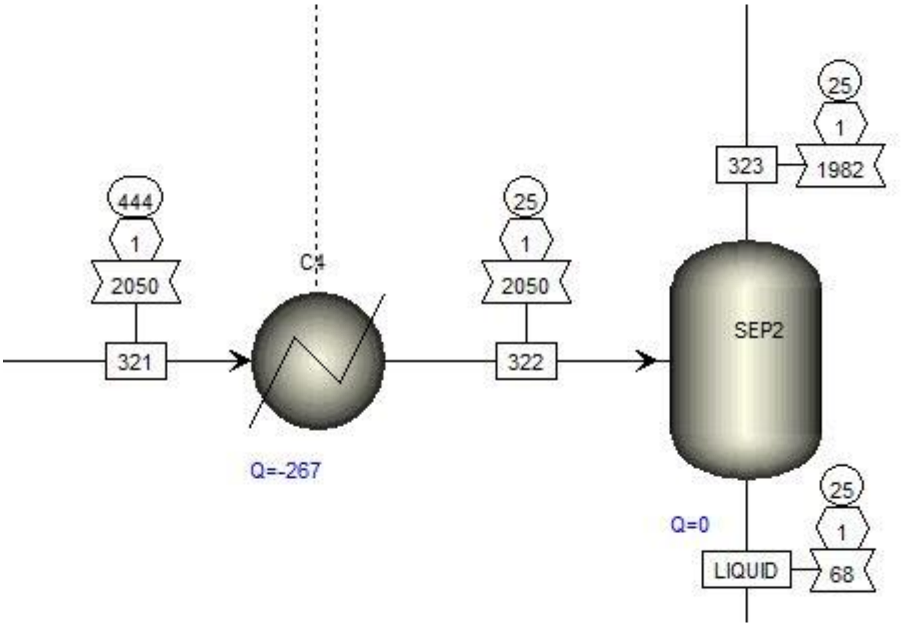


Fig 56: Pyrolysis Oil extraction unit

Reformer and heat recovery unit

The Reformer unit is modelled as an RGibbs reactor with a specified reactor temperature of 850°C. The heat requirement for this reactor is satisfied from the heat energy coming from the combustor unit. The outlet gases from the reactor are cooled at 300°C by a HEATER block (Cooler 1). The energy from this block is used to heat the gases after SHIFT reactor to meet the energy requirements of CALCINATOR reactor. The details of this unit are presented in **Fig 57.** The steam requirement for this unit is adjusted in such a way so that most of the Methane and Ethane is converted into CO and H_2.

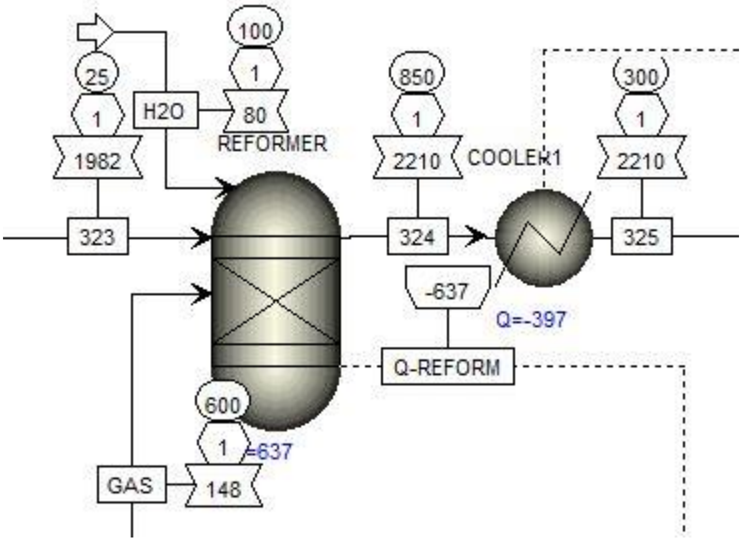


Fig: 57 Details of Reformer unit

Tar cracking unit

The tar cracking unit is modelled as two RGibbs reactors (Fe_2O_3 looping system) presented in **Fig 58**. 5% of the Fe_2O_3 was considered as waste. The feed of pure Fe_2O_3 was adjusted in such a way that SOLID Carbon was not generated in the recycle stream (SOLID-STREAM). The Oxygen requirement for the OXIDIZER was calculated using a calculator block. The OXIDIZER unit operates at 900°C. Most of the tar is cracked in the looping system and then joined the Reformer reactor for reforming of the available Methane and Ethane.

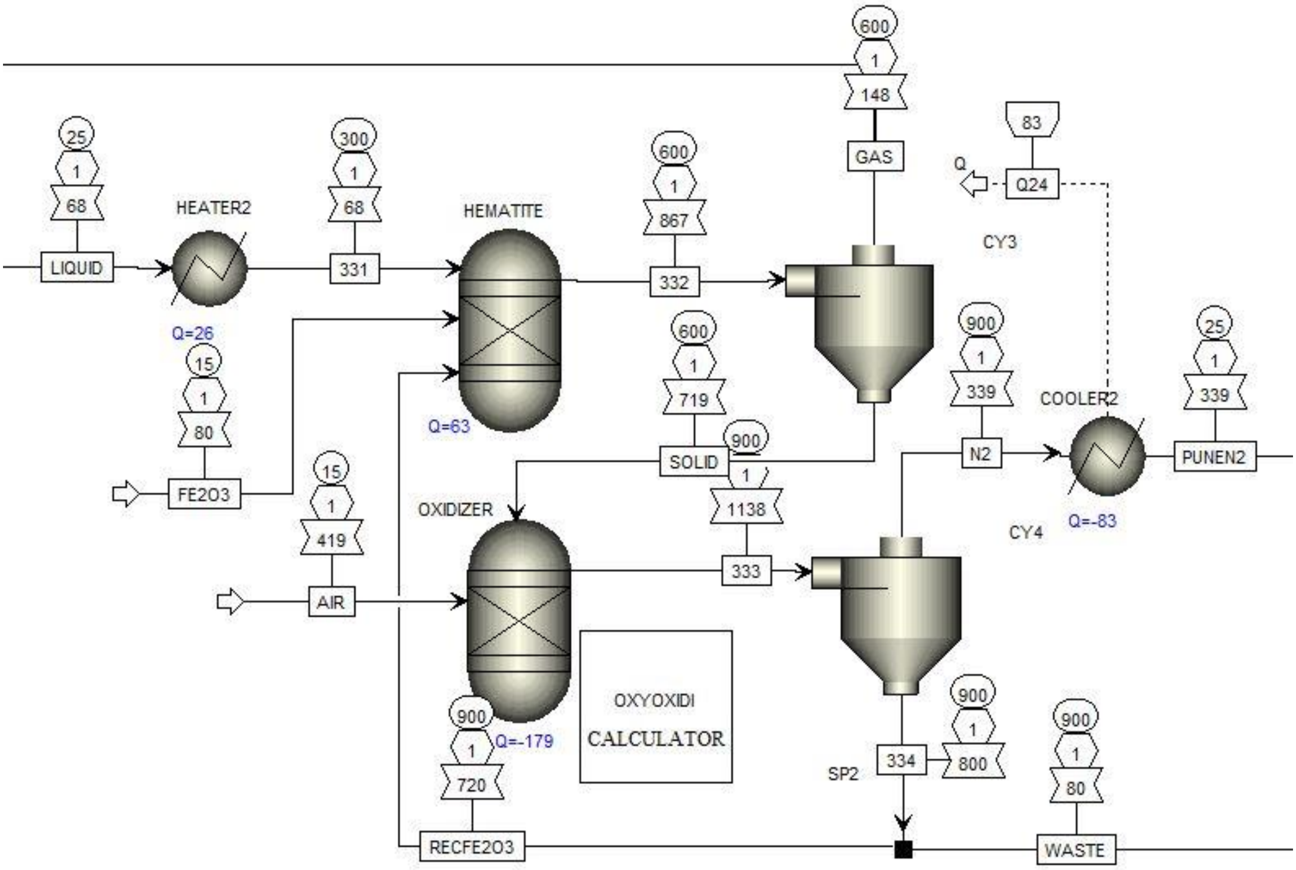
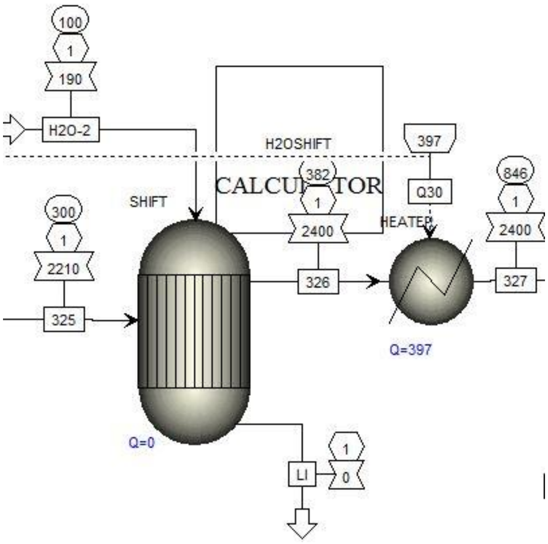
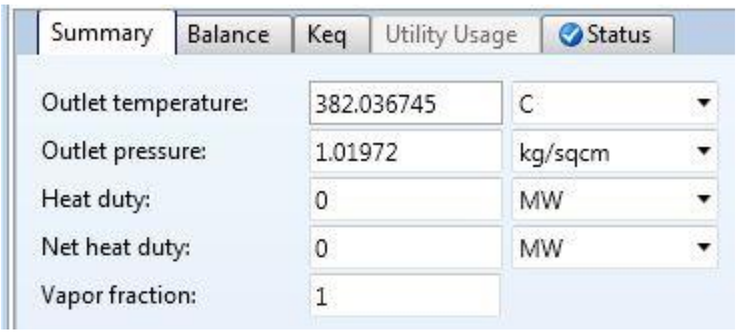


Fig 58: Tar cracking unit

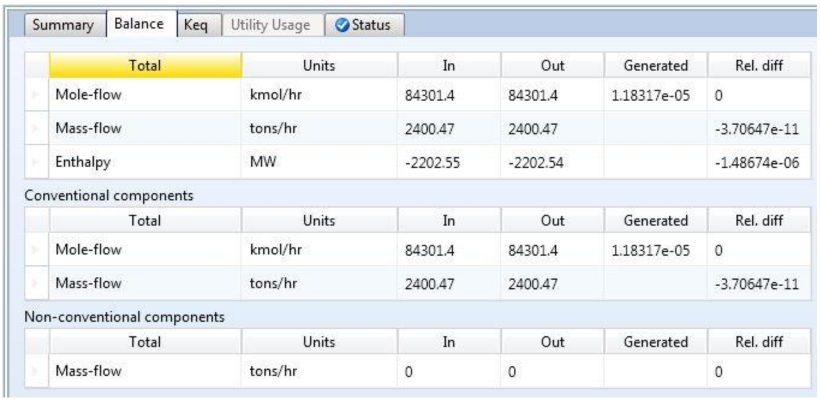
SHIFT reactor and heat recovery unit

The SHIFT reactor is modelled as a REquil reactor with duty 0 (So that process model will calculate outlet temperature in phase and reaction equilibrium). The steam requirement for this unit is calculated through a calculator block.





Outlet temperature:	382.036745	C
Outlet pressure:	1.01972	kg/sqcm
Heat duty:	0	MW
Net heat duty:	0	MW
Vapor fraction:	1	



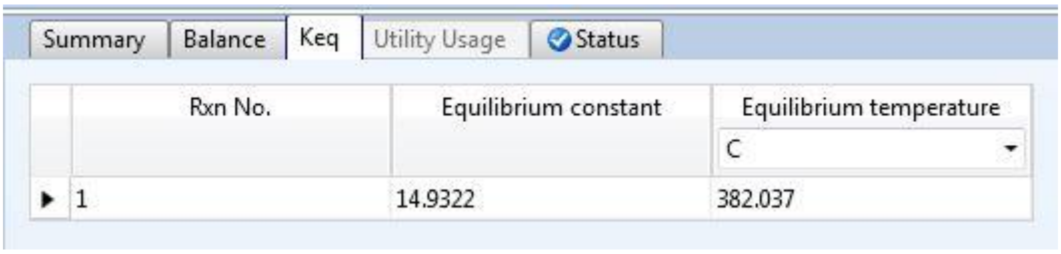
Total	Units	In	Out	Generated	Rel. diff
Mole-flow	kmol/hr	84301.4	84301.4	1.18317e-05	0
Mass-flow	tons/hr	2400.47	2400.47		-3.70647e-11
Enthalpy	MW	-2202.55	-2202.54		-1.48674e-06

Conventional components

Total	Units	In	Out	Generated	Rel. diff
Mole-flow	kmol/hr	84301.4	84301.4	1.18317e-05	0
Mass-flow	tons/hr	2400.47	2400.47		-3.70647e-11

Non-conventional components

Total	Units	In	Out	Generated	Rel. diff
Mass-flow	tons/hr	0	0		0



Rxn No.	Equilibrium constant	Equilibrium temperature C
1	14.9322	382.037

Fig 59: Details of Shift reactor unit

Gas cleaning and CO_2 separation unit

The gas cleaning and CO_2 separation unit was modelled as a $CACO_3$ looping unit presented in **Fig 60**. An extensive study was carried out, and we choose this option. $CACO_3$ looping can meet dual purposes, the pollutant removal as well as the separation of CO_2 from the mainstream. The main advantage of this unit is it does not require fresh water stream as like Ammine scrubbing. There is two RGibbs reactors with two cyclone separator. In the carbonator reactor, the reaction enthalpy is exothermic and specified as 650°C. The stream of gases 327 enters the reactor and reacts with CAO.

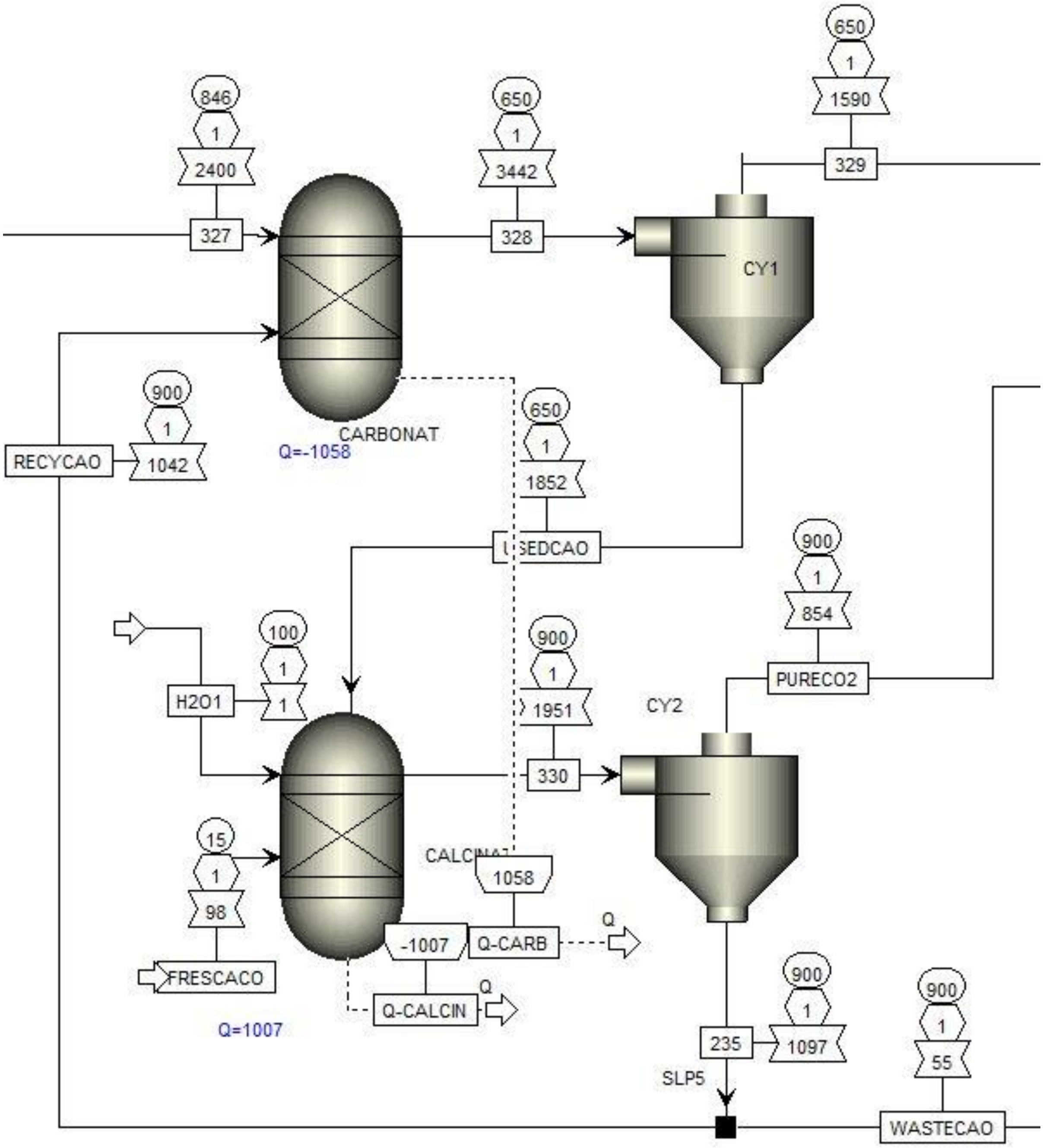


Fig 60: $CACO_3$ looping unit for gas cleaning and CO_2 separation

The reaction enthalpy in the CALCINATOR is endothermic. The heat requirement of this block is satisfied by the excess heat energy from the CARBONATOR unit. The CARBONATOR block will have excess energy to run the CALCINATOR if efficient heat removal is done before the REFORMER reactor that is used to heat the gases before CARBONATOR.

5% of the CAO was considered as waste. The entire unit was modelled in such a way so that in the recycle stream (USED CAO) there is no CAO going to CALCINATOR unit. A flow of steam (H2O) was added to the CALCINATOR so that it can drive the CO_2 and other pollutant gases from the system.

Heat recovery and PSA for H2 Separation

Energy is recovered from the clean gas and pure CO_2 stream through HEATER block as presented in **Fig 61**. The energy is then used for steam generation for the dryer. The H2 separation unit is considered as a component splitter.

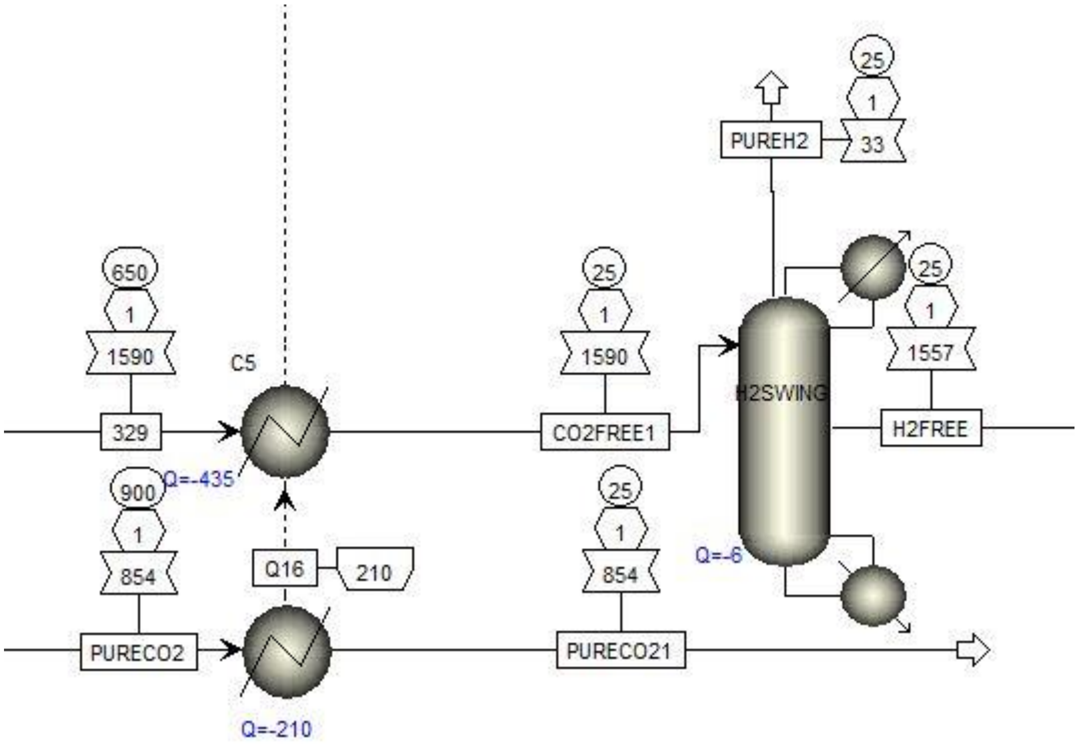


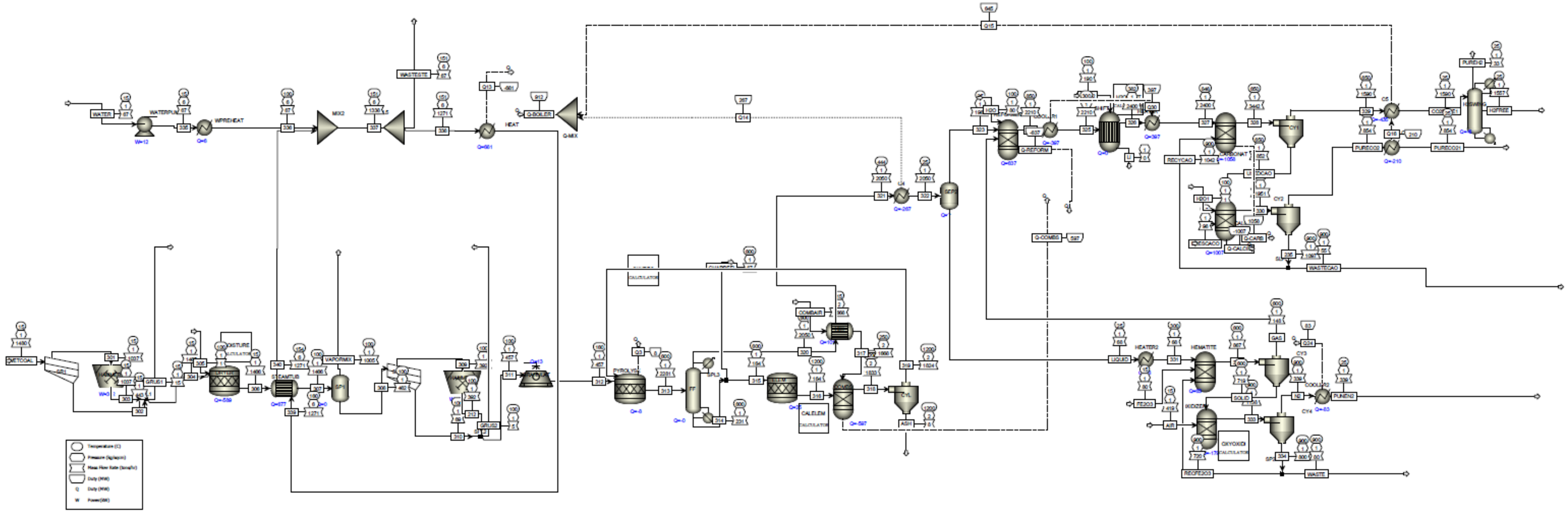
Fig 61: Heat Recovery and H2 Separation unit

Table 4 present the stream results

Table 4

		WET COAL	STEAM	DRY-COAL	CAO	FE2O3	PURE-CO2	PURE-H2
Mass Flow	TONS/HR	1480	67	456.9228	98	80	854.4946	32.61692
Mass Enthalpy	MW	-3951.865	-268.5413	-458.5888	-298.108	-104.231	-1711.653	1.52E-14
Mass flow rate								
H2	TONS/HR	0	0	0	0	0	0.000378906	32.61692
CO	TONS/HR	0	0	0	0	0	2.311444	0
CO2	TONS/HR	0	0	0	0	0	849.4199	0
SO2	TONS/HR	0	0	0	0	0	1.766241	0
SO3	TONS/HR	0	0	0	0	0	1.48E-06	0
NO2	TONS/HR	0	0	0	0	0	0	0
NO	TONS/HR	0	0	0	0	0	0	0
S	TONS/HR	0	0	0	0	0	0	0
H2O	TONS/HR	0	67	0	0	0	0.9965985	0
C	TONS/HR	0	0	0	0	0	0	0
C2H6	TONS/HR	0	0	0	0	0	3.38E-32	0
CH4	TONS/HR	0	0	0	0	0	2.84E-16	0
H2S	TONS/HR	0	0	0	0	0	2.89E-05	0
C3H8	TONS/HR	0	0	0	0	0	0	0
C3H6-2	TONS/HR	0	0	0	0	0	0	0
O2	TONS/HR	0	0	0	0	0	2.81E-09	0
N2	TONS/HR	0	0	0	0	0	0	0
NH3	TONS/HR	0	0	0	0	0	0	0
C10H8	TONS/HR	0	0	0	0	0	0	0
C6H6O	TONS/HR	0	0	0	0	0	0	0
TERT--01	TONS/HR	0	0	0	0	0	0	0
C2H4	TONS/HR	0	0	0	0	0	1.55E-26	0
P-CRE-01	TONS/HR	0	0	0	0	0	0	0
CAO	TONS/HR	0	0	0	0	0	0	0

CACO3	TONS/HR	0	0	0	98	0	0	0
CASO3	TONS/HR	0	0	0	0	0	0	0
CASO4	TONS/HR	0	0	0	0	0	0	0
CAS	TONS/HR	0	0	0	0	0	0	0
FE	TONS/HR	0	0	0	0	0	0	0
FE2O3	TONS/HR	0	0	0	0	80	0	0
FEO	TONS/HR	0	0	0	0	0	0	0
FE3O4	TONS/HR	0	0	0	0	0	0	0
COAL	TONS/HR	1480	0	456.9228	0	0	0	0
CHAR	TONS/HR	0	0	0	0	0	0	0
ASH	TONS/HR	0	0	0	0	0	0	0


WATER
WATERPUMP
W=12
WPREHEAT
Q=6
MIX2
WASTESTE
Q13
-901
HEAT
Q=901
Q-BOILER
912
Q-MIX
Q15
645
Q14
267
WETCOAL
SR1
GRUS1
MOISTURE CALCULATOR
DRYER
Q=-589
STEAMTUB
Q=877
VAPORMIX
SP1
GRUS2
Q=13
PYROLYSI
Q3
FF
SPL3
CALCULATOR
COMBAIR
ELEM
CALELEM CALCULATOR
COMBS
Q=-597
CYL
ASH
Q-COMBS
597
SEP2
Q=-267
H2O
REFORM
COOLER1
Q-REFORM
Q=637
Q=-397
SHIFT
Q30
CARBONAT
RECYCAO
USEDCAO
H2O1
DESCACO
Q-CARB
Q-CALCIN
Q=1007
CY1
CY2
WASTECAO
C5
Q16
210
PURECO2
PURECO21
Q=-210
PURIH2
H2SWING
H2FREE
LIQUID
HEATER2
HEMATITE
GAS
Q24
83
CY3
SOLID
COOLER2
Q=-83
PURIN2
FE2O3
OXIDIZER
OXYOXIDI CALCULATOR
AIR
CY4
SP2
RECFE2O3
WASTE
Temperature (C)
Pressure (kg/sqcm)
Mass Flow Rate (tons/hr)
Duty (MW)
Q Duty (MW)
W Power(kW)

4. Detailed simulation results of Option-4

This study presents a mass-energy balance for H_2 production plant using yallourn coal through gasification. The production target of H_2 is 770 tons/day.

The entire plant has several sections from feed coal preparation to gasification as well as downstream gas cleaning followed by pressure-swing adsorption of H_2 separation.

Several assumptions were made during simulation. The simulation begins with the mass balance from dry coal to gasified products. The gasifier runs at 30 bar pressure. So it is essential to correct the 1 bar coal pyrolysis data for 30 bar pressure as presented in **Table 5**.

Table 5			
PRESSURE	1	PRESSURE	30
GAS YIELD	42	GAS YIELD	32.57
SOLID YEILD	58	SOLID YEILD	67.43
COMPOSITION	1000°C	COMPOSITION	1000°C
CO	44	CO	24.55
H2	42	H2	1.69
CO2	2.83	CO2	2.48
CH4	3.86	CH4	1.23
C2H6	0	C2H6	0.00
H2O	7.31	H2O	2.62
TOTAL	100	CHAR	67.43
		TOTAL	100

BAR	AT GAS YIELD									
AT 30 BAR	(g)	AT 100%	mol%	C	H	N	S	O	ASH	TOTAL
CO	24.55	0.25	0.01	10.53	0.00	0.00	0.00	14.02	0.00	24.55
H2	1.69	0.02	0.01	0.00	1.69	0.00	0.00	0.00	0.00	1.69
CO2	2.48	0.02	0.00	0.68	0.00	0.00	0.00	1.80	0.00	2.48
CH4	1.23	0.01	0.00	0.92	0.31	0.00	0.00	0.00	0.00	1.23
C2H6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H2O	2.62	0.03	0.00	0.00	0.29	0.00	0.00	2.33	0.00	2.62
CHAR	67.43	0.67	0.67	55.89	2.36	0.45	0.29	5.91	2.53	67.43

TOTAL	100.00	1.00	0.69	68.01	4.65	0.45	0.29	24.07	2.53	100.00

The parent coal was considered with 60% moisture content. The moisture content reduced to 2.7% after steam tube dryer.

The details of the individual unit operation are as follows:

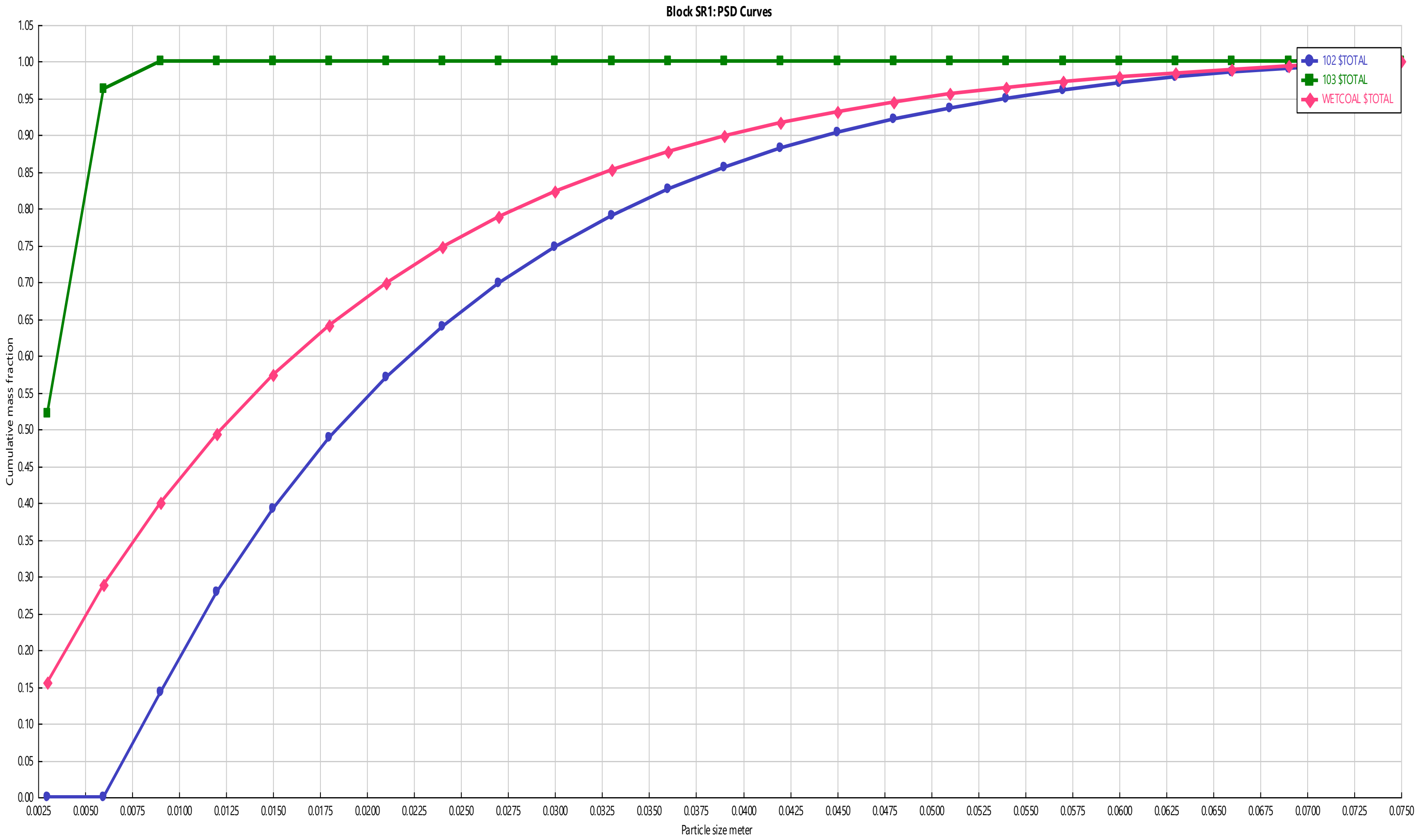
Coal storage

Coal storage is not a process operation and not modelled.

Primary screening and Hammering

The raw coal is considered with 60% moisture with a PSD as described before. This wet coal goes to a primary hammer mill followed by a preliminary screening.

The primary screener is modelled as default SCREEN with a cut size of 0.0063 m (Coarse-split) to meet the primary Hammer mill requirement. The separation stage and efficiency is presented in **Fig 62**.


Block SR1: PSD Curves
Cumulative mass fraction
1.05
1.00
0.95
0.90
0.85
0.80
0.75
0.70
0.65
0.60
0.55
0.50
0.45
0.40
0.35
0.30
0.25
0.20
0.15
0.10
0.05
0.00
0.0025
0.0050
0.0075
0.0100
0.0125
0.0150
0.0175
0.0200
0.0225
0.0250
0.0275
0.0300
0.0325
0.0350
0.0375
0.0400
0.0425
0.0450
0.0475
0.0500
0.0525
0.0550
0.0575
0.0600
0.0625
0.0650
0.0675
0.0700
0.0725
0.0750
Particle size meter
102 $TOTAL
103 $TOTAL
WETCOAL $TOTAL

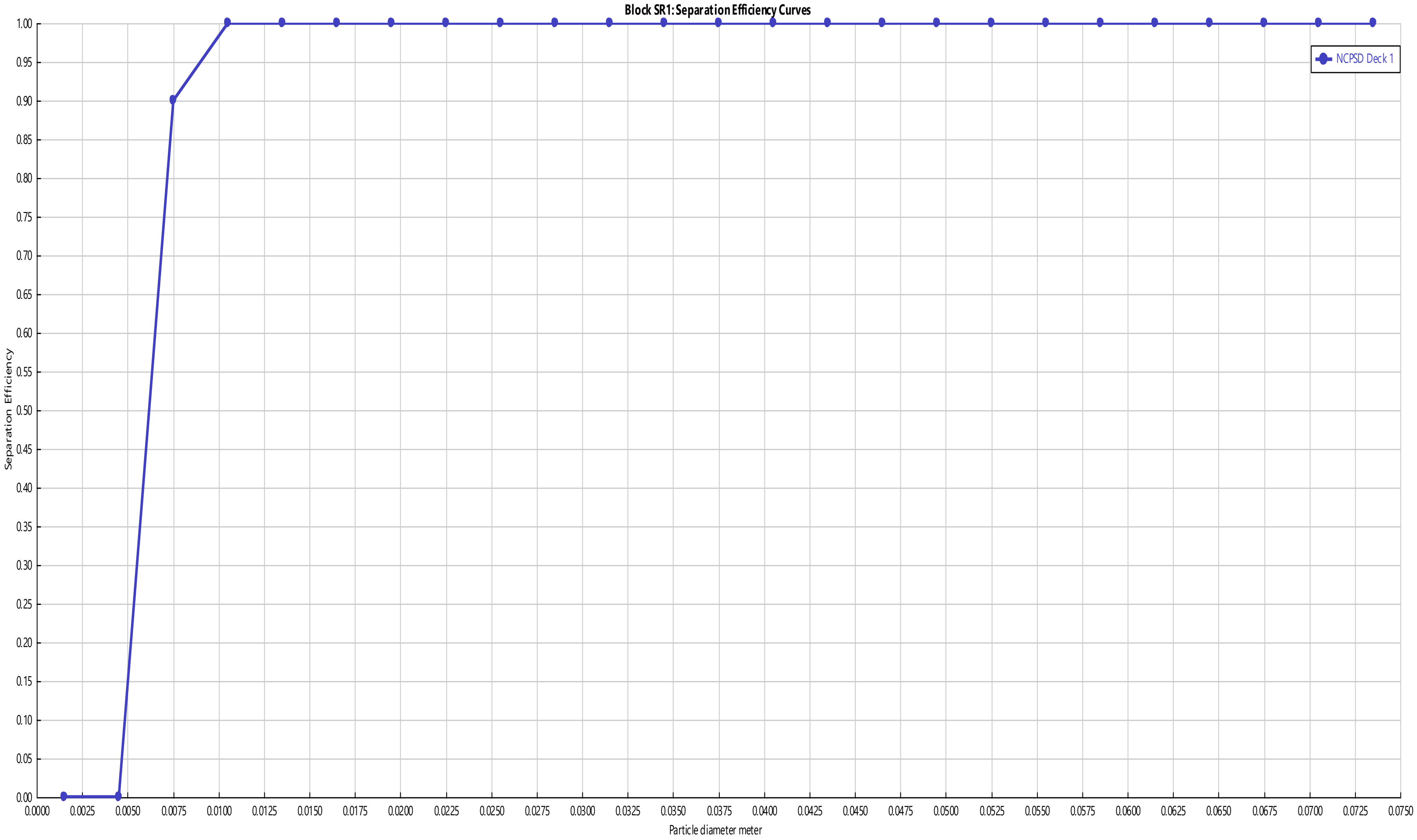
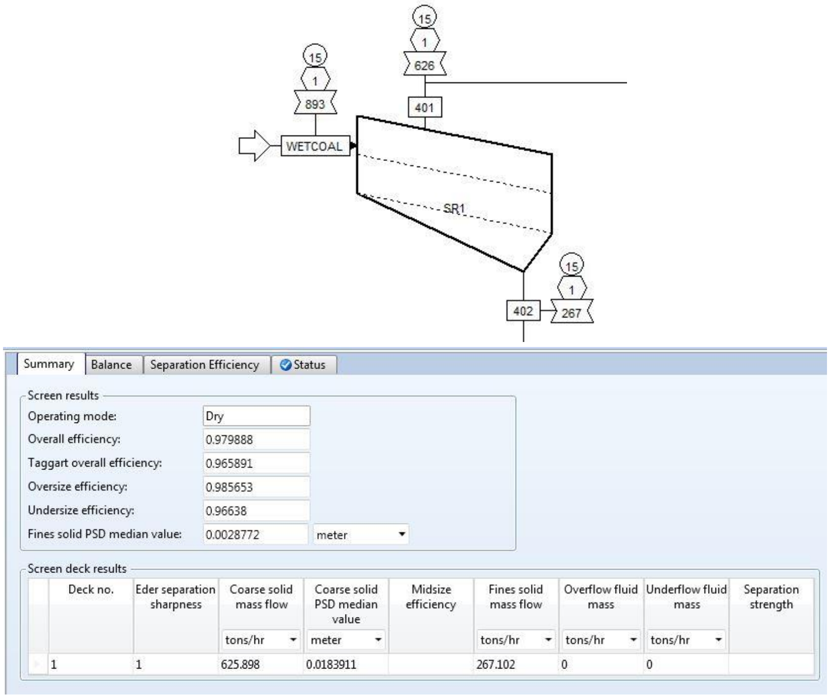


Fig 62: Primary screening separation and efficiency

The details of the primary SCREENING machine is presented in **Fig 63**



Screen results

Operating mode: Dry
Overall efficiency: 0.979888
Taggart overall efficiency: 0.965891
Oversize efficiency: 0.985653
Undersize efficiency: 0.96638
Fines solid PSD median value: 0.0028772 meter

Screen deck results

Deck no.	Eder separation sharpness	Coarse solid mass flow	Coarse solid PSD median value	Midsize efficiency	Fines solid mass flow	Overflow fluid mass	Underflow fluid mass	Separation strength
		tons/hr	meter		tons/hr	tons/hr	tons/hr	
1	1	625.898	0.0183911		267.102	0	0	

Fig 63: Details of primary screener

The hammer mill is considered as a HAMMER CRUSHER with default in the built model. The PSD curve shape, i.e. RRSB distribution function using dispersion parameter=2, D63 particle size adjusted to suit ~98 wt% particles <6 mm and secondary split. The open cut method with Bands law (HG-index 130) is used.

The PSD after primary hammering is presented in **Fig 64.**

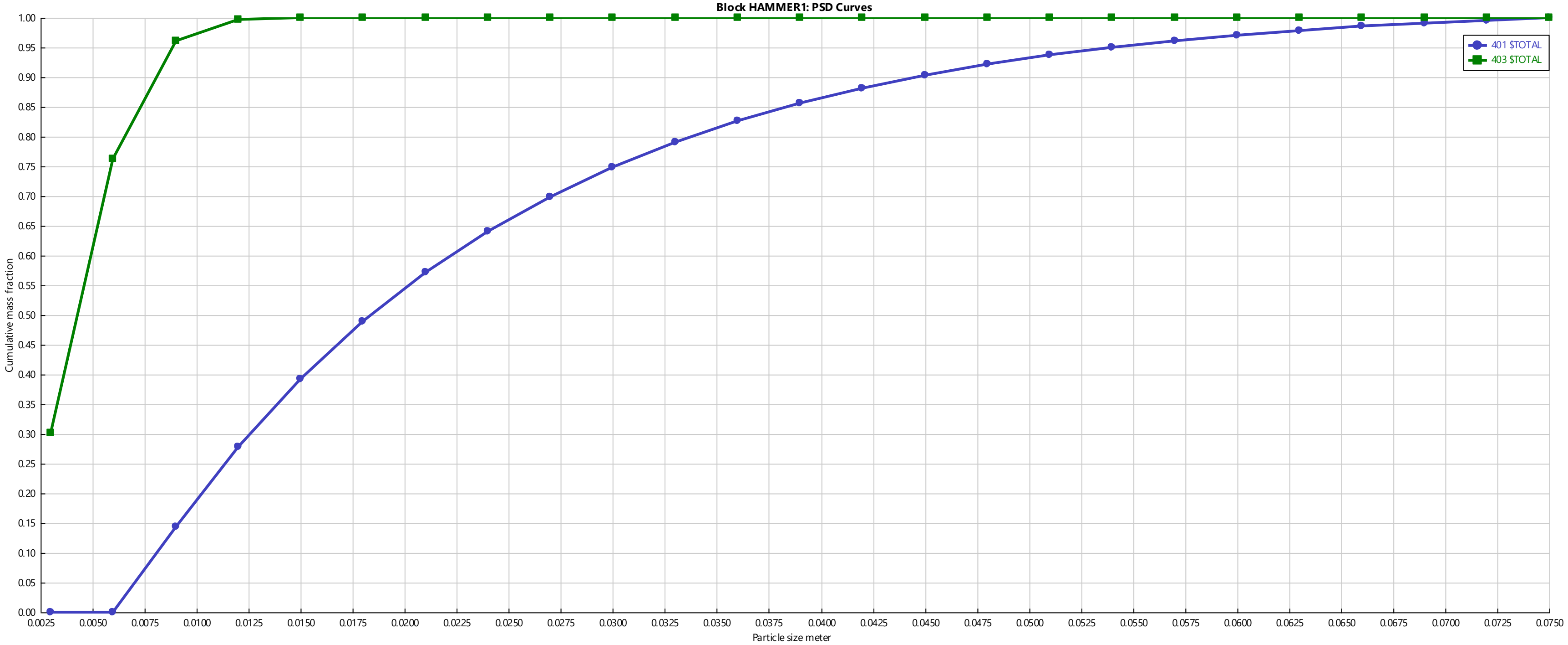
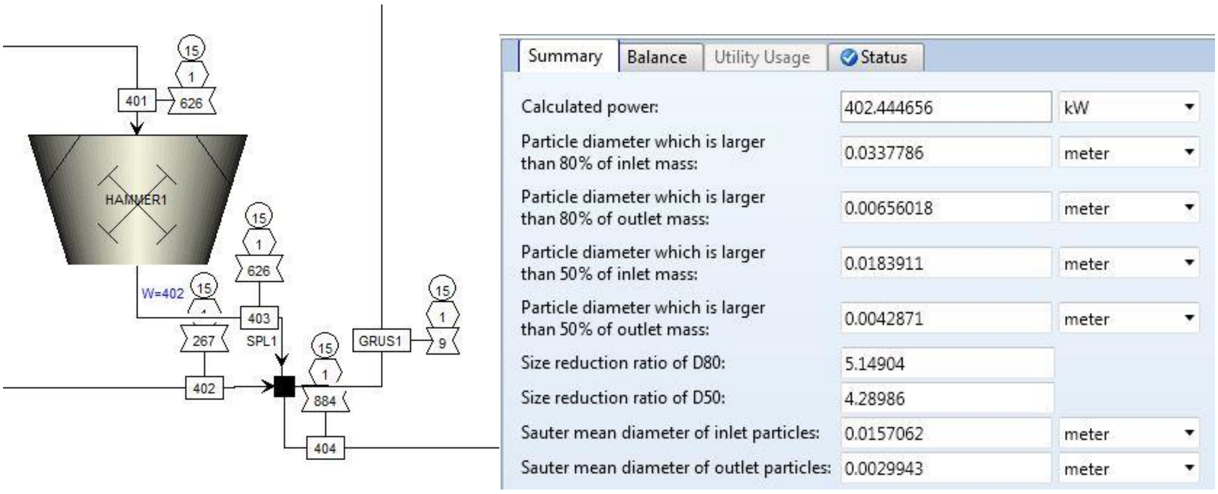


Fig 64: PSD after primary hammering

The details of the primary Hammer machine is presented in **Fig 65**



Calculated power:	402.444656	kW
Particle diameter which is larger than 80% of inlet mass:	0.0337786	meter
Particle diameter which is larger than 80% of outlet mass:	0.00656018	meter
Particle diameter which is larger than 50% of inlet mass:	0.0183911	meter
Particle diameter which is larger than 50% of outlet mass:	0.0042871	meter
Size reduction ratio of D80:	5.14904	
Size reduction ratio of D50:	4.28986	
Sauter mean diameter of inlet particles:	0.0157062	meter
Sauter mean diameter of outlet particles:	0.0029943	meter

Fig 65: Details of Primary Hammer

**1% of raw coal was considered as grus and was not accounted in the mainstream. An FSPLIT was used for this purpose.

Steam Tube dryer

The steam tube dryer is modelled with 3 different default inbuilt unit operation unit. An RStoic is used to convert the non-conventional moisture present in coal to convert conventional pure water at 15°C based on Eq (1):

WETCOAL--> 0.0555084 H_2O …….. (1)

The pure conventional water is passed through a heat exchanger (HX) with a specification of vapour fraction 1 at the outlet of the cold fluid. All the water that is vaporised in the HX unit is separate in a splitter (SP1). **1% of entrainment of coal was considered in the water vapour.** The overall unit of the dryer is presented in **Fig 66.**

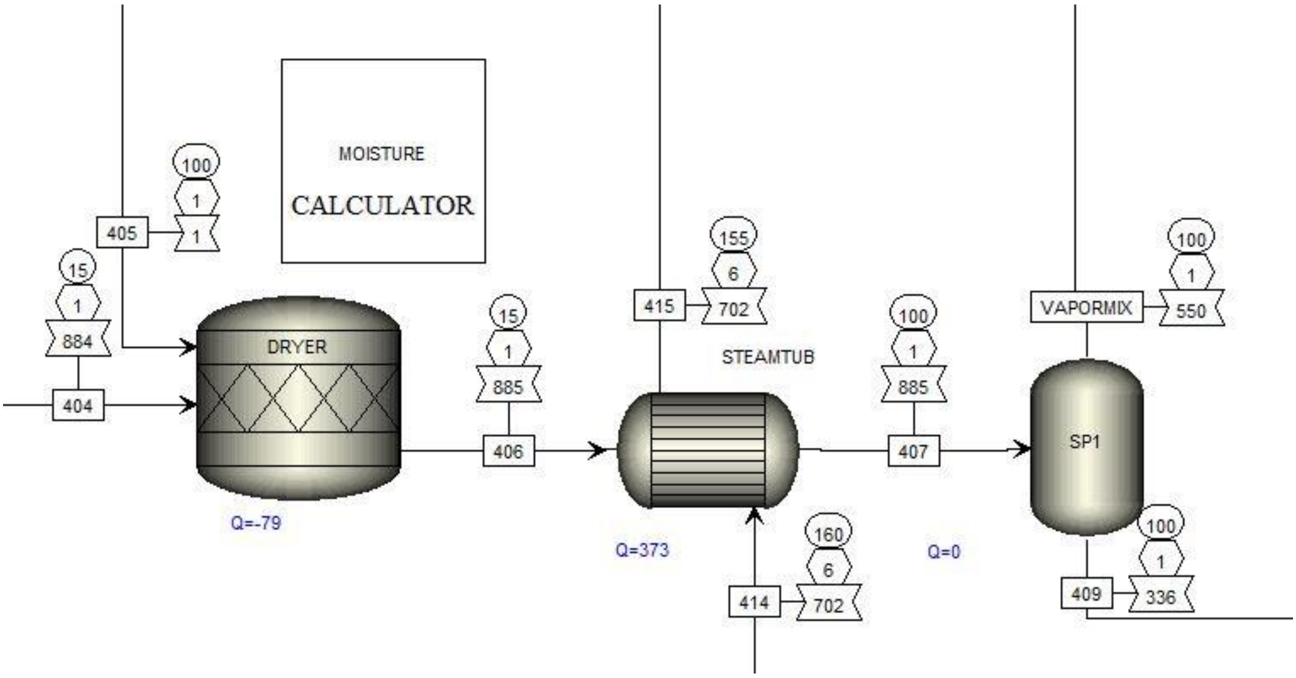
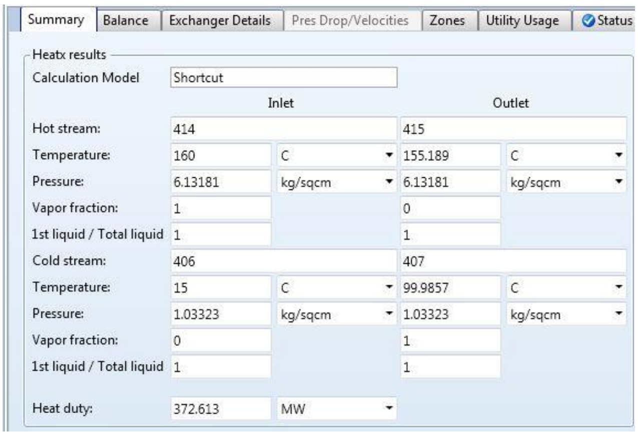


Fig 66: Steam tube dryer

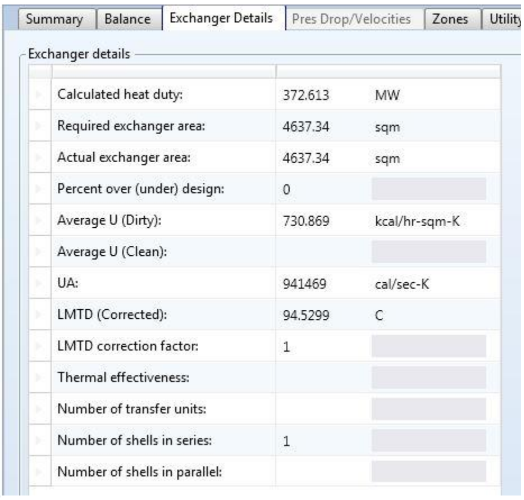
A stream of air 405 is introduced in the DRYER to pull out the moisture from the tubes. 160°C and 5 barg steam were used in the hot stream of the HX. The steam requirement was adjusted such a way that the temperature drop after HX of the hot stream is 155°C. The details of the HX unit is presented in **Fig 67.**



Heatx results

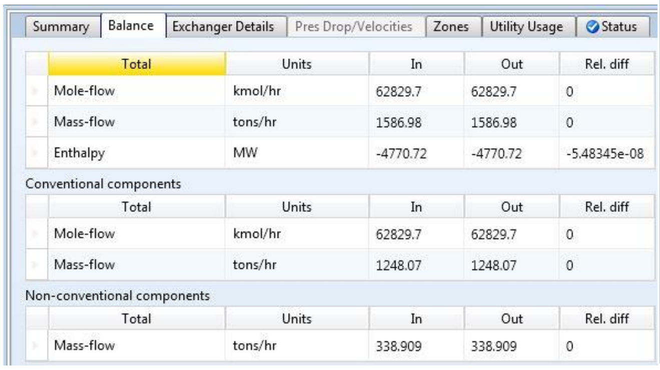
Calculation Model: Shortcut

	Inlet		Outlet	
Hot stream:	414		415	
Temperature:	160	C	155.189	C
Pressure:	6.13181	kg/sqcm	6.13181	kg/sqcm
Vapor fraction:	1		0	
1st liquid / Total liquid	1		1	
Cold stream:	406		407	
Temperature:	15	C	99.9857	C
Pressure:	1.03323	kg/sqcm	1.03323	kg/sqcm
Vapor fraction:	0		1	
1st liquid / Total liquid	1		1	
Heat duty:	372.613	MW		



Exchanger details

Calculated heat duty:	372.613	MW
Required exchanger area:	4637.34	sqm
Actual exchanger area:	4637.34	sqm
Percent over (under) design:	0	
Average U (Dirty):	730.869	kcal/hr-sqm-K
Average U (Clean):		
UA:	941469	cal/sec-K
LMTD (Corrected):	94.5299	C
LMTD correction factor:	1	
Thermal effectiveness:		
Number of transfer units:		
Number of shells in series:	1	
Number of shells in parallel:		



Total	Units	In	Out	Rel. diff
Mole-flow	kmol/hr	62829.7	62829.7	0
Mass-flow	tons/hr	1586.98	1586.98	0
Enthalpy	MW	-4770.72	-4770.72	-5.48345e-08

Conventional components

Total	Units	In	Out	Rel. diff
Mole-flow	kmol/hr	62829.7	62829.7	0
Mass-flow	tons/hr	1248.07	1248.07	0

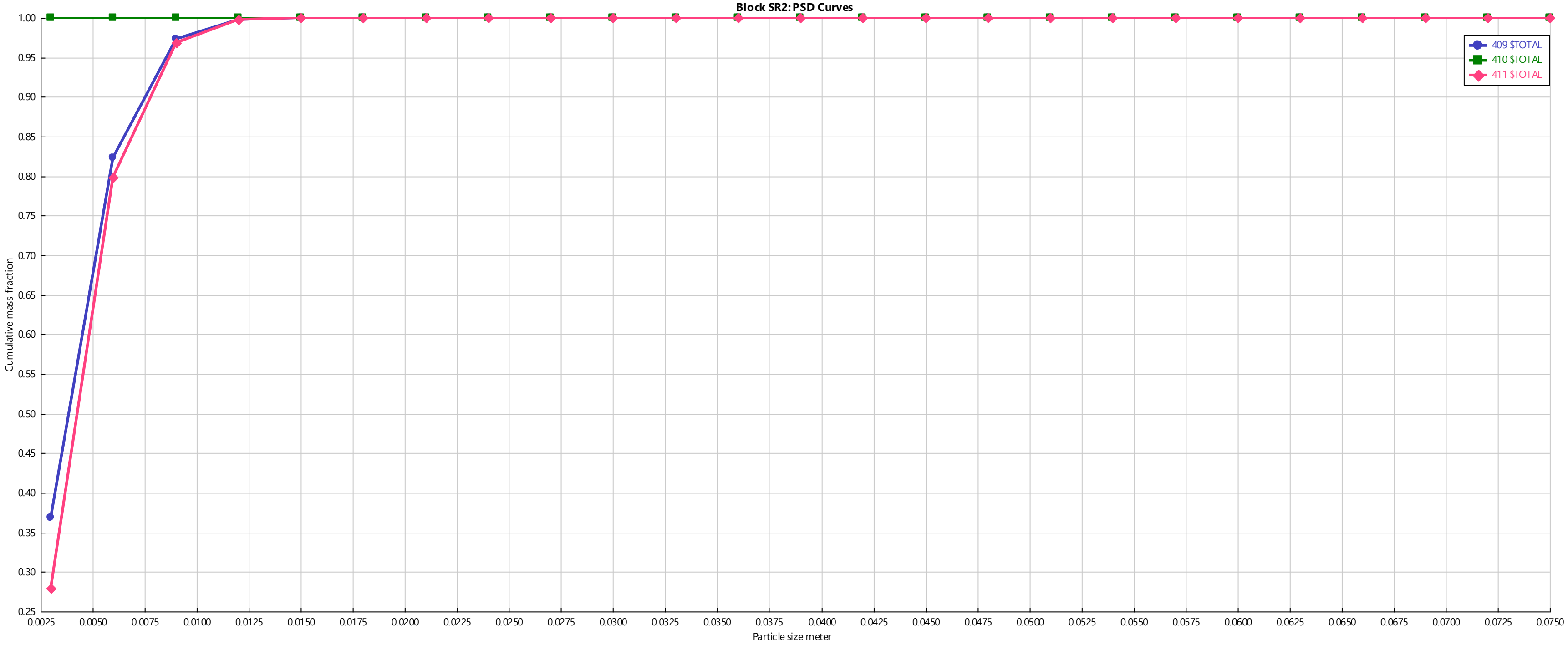
Non-conventional components

Total	Units	In	Out	Rel. diff
Mass-flow	tons/hr	338.909	338.909	0

Fig 67: Details of Steam tube dryer unit

Secondary screening and Hammering

The dry coal goes to a secondary screening and hammering for further refinement. The secondary screener is modelled as default SCREEN with a cut size of 0.001 m (Coarse-split) to meet the secondary Hammer mill requirement. The separation stage and efficiency is presented in **Fig 68**.


Block SR2: PSD Curves
Cumulative mass fraction
1.00
0.95
0.90
0.85
0.80
0.75
0.70
0.65
0.60
0.55
0.50
0.45
0.40
0.35
0.30
0.25
0.0025
0.0050
0.0075
0.0100
0.0125
0.0150
0.0175
0.0200
0.0225
0.0250
0.0275
0.0300
0.0325
0.0350
0.0375
0.0400
0.0425
0.0450
0.0475
0.0500
0.0525
0.0550
0.0575
0.0600
0.0625
0.0650
0.0675
0.0700
0.0725
0.0750
Particle size meter
409 $TOTAL
410 $TOTAL
411 $TOTAL

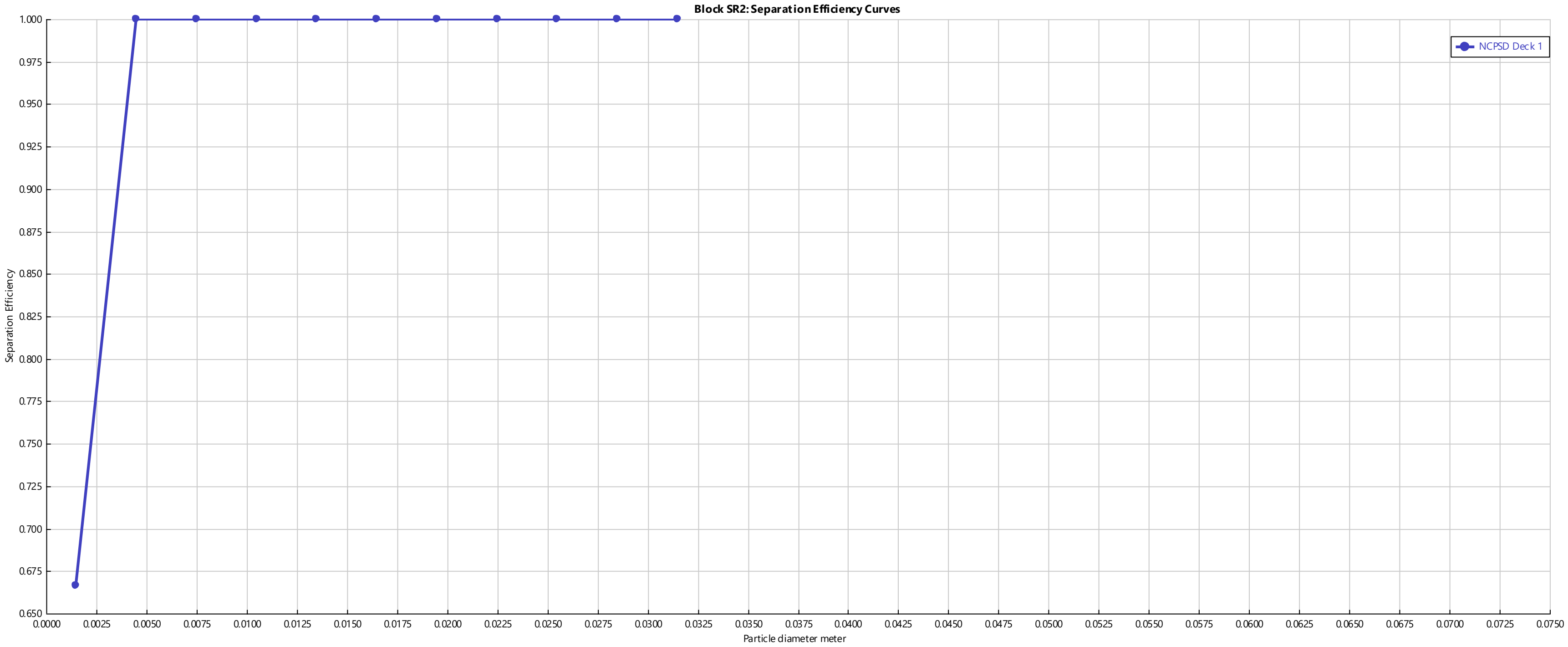
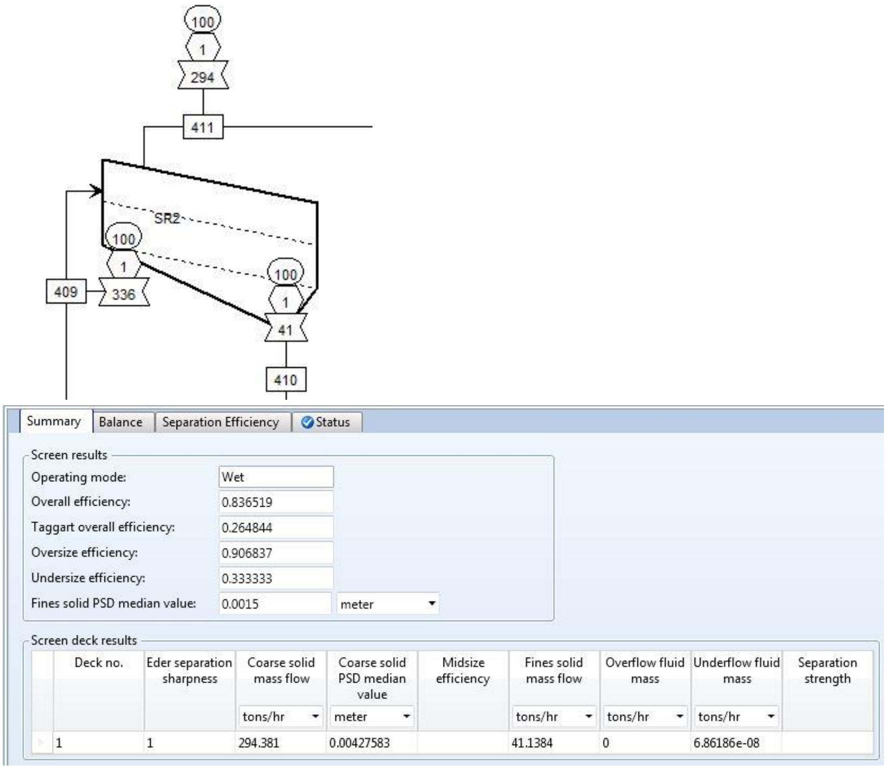


Fig 68: Secondary screening separation and efficiency

The details of the primary SCREENING machine is presented in **Fig 69**



Screen results

Operating mode: Wet
Overall efficiency: 0.836519
Taggart overall efficiency: 0.264844
Oversize efficiency: 0.906837
Undersize efficiency: 0.333333
Fines solid PSD median value: 0.0015 meter

Screen deck results

Deck no.	Eder separation sharpness	Coarse solid mass flow tons/hr	Coarse solid PSD median value meter	Midsize efficiency	Fines solid mass flow tons/hr	Overflow fluid mass tons/hr	Underflow fluid mass tons/hr	Separation strength
1	1	294.381	0.00427583		41.1384	0	6.86186e-08	

Fig 69: Details of secondary screener

The secondary hammer mill is considered as a HAMMER CRUSHER with default inbuilt model. The PSD curve shape i.e., RRSB distribution function using dispersion parameter=2, D63 particle size adjusted to suit ~98 wt% particles <1mm and secondary split.

The PSD after primary hammering is presented in **Fig 70.**

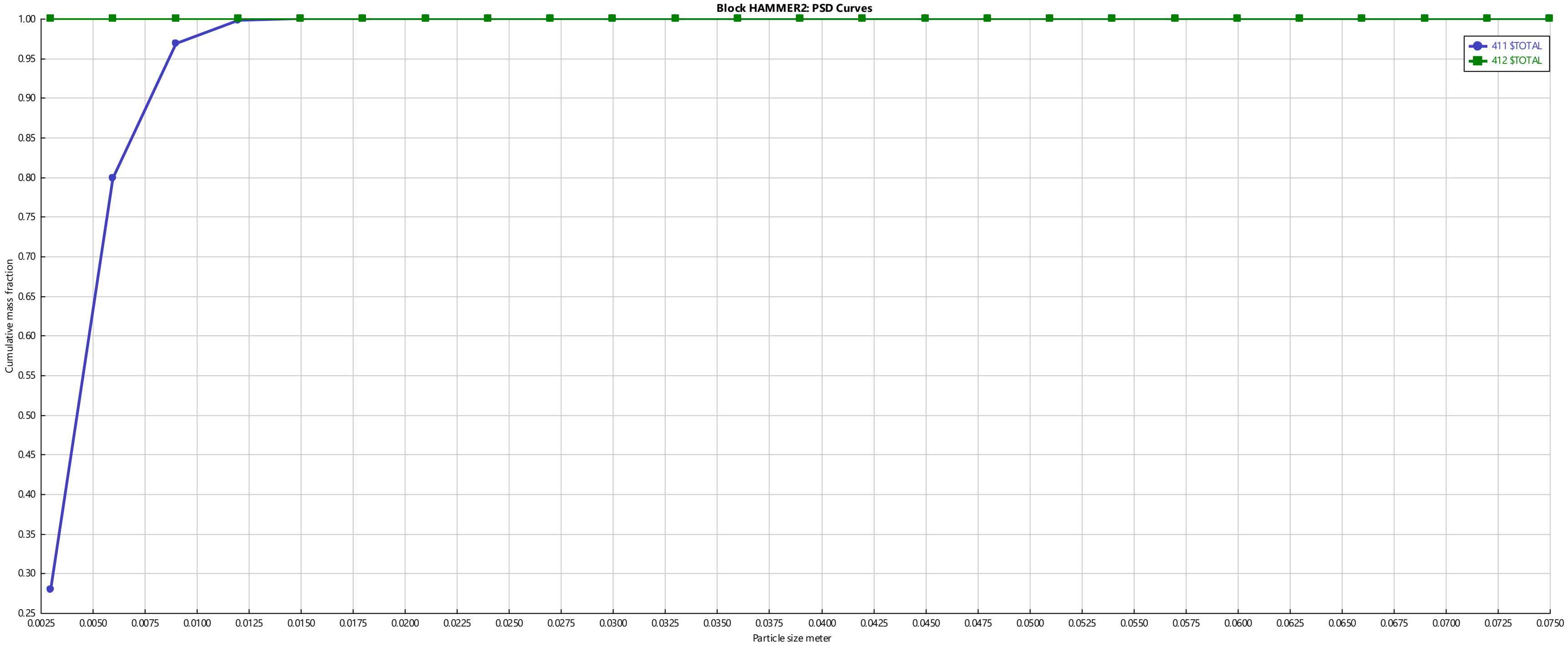
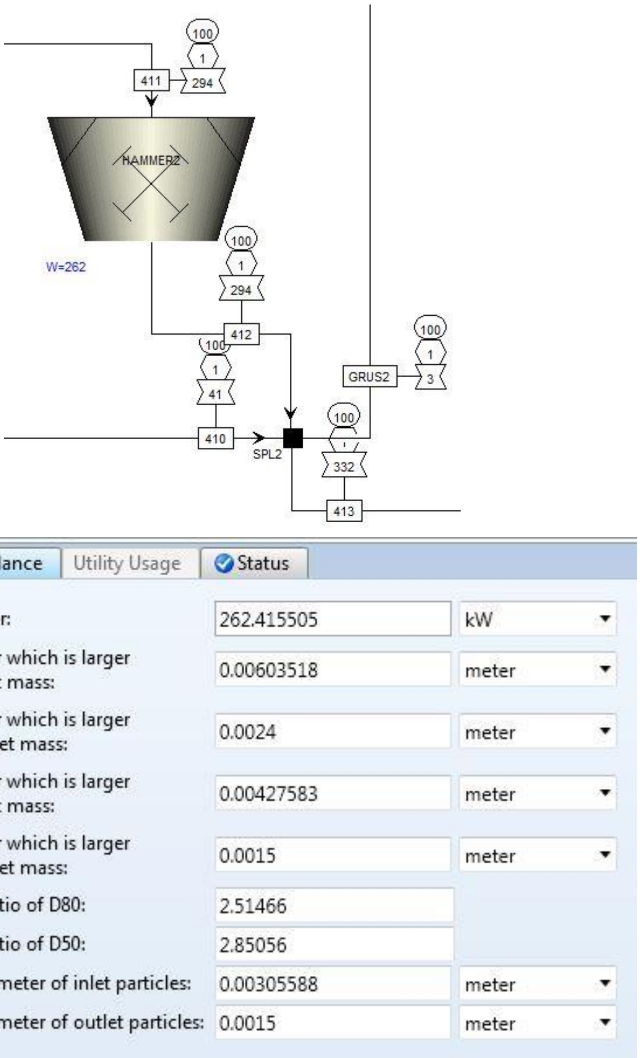


Fig 70: PSD of Secondary Hammer

The details of the secondary Hammer machine is presented in **Fig 71**



Calculated power: 262.415505 kW

Particle diameter which is larger than 80% of inlet mass: 0.00603518 meter

Particle diameter which is larger than 80% of outlet mass: 0.0024 meter

Particle diameter which is larger than 50% of inlet mass: 0.00427583 meter

Particle diameter which is larger than 50% of outlet mass: 0.0015 meter

Size reduction ratio of D80: 2.51466

Size reduction ratio of D50: 2.85056

Sauter mean diameter of inlet particles: 0.00305588 meter

Sauter mean diameter of outlet particles: 0.0015 meter

Fig 71: Details of Secondary Hammer

**1% of raw coal was considered as grus and was not accounted in the mainstream. An FSPLIT was used for this purpose.

Gasification unit

The gasification unit is considered as two-stage Oxy-fired entrained flow gasifier operated at 1400°C and 30 bar pressure presented in **Fig 72**. The dry coal is split into two streams. Stream 1 has 15% of the dry cola that goes to combustion unit (1st stage). The dry coal is devolatilized in an RStoic reactor and pressure correction is carried out in the volatile gases. The char is then decomposed to its elementary composition (RStoic) and enter the Combustor unit (RGibbs). The air requirement for the combustor unit is carried out with a calculator block considering 20% excess oxygen to sustain an entrained condition in the second gasifier for at least 6 seconds.

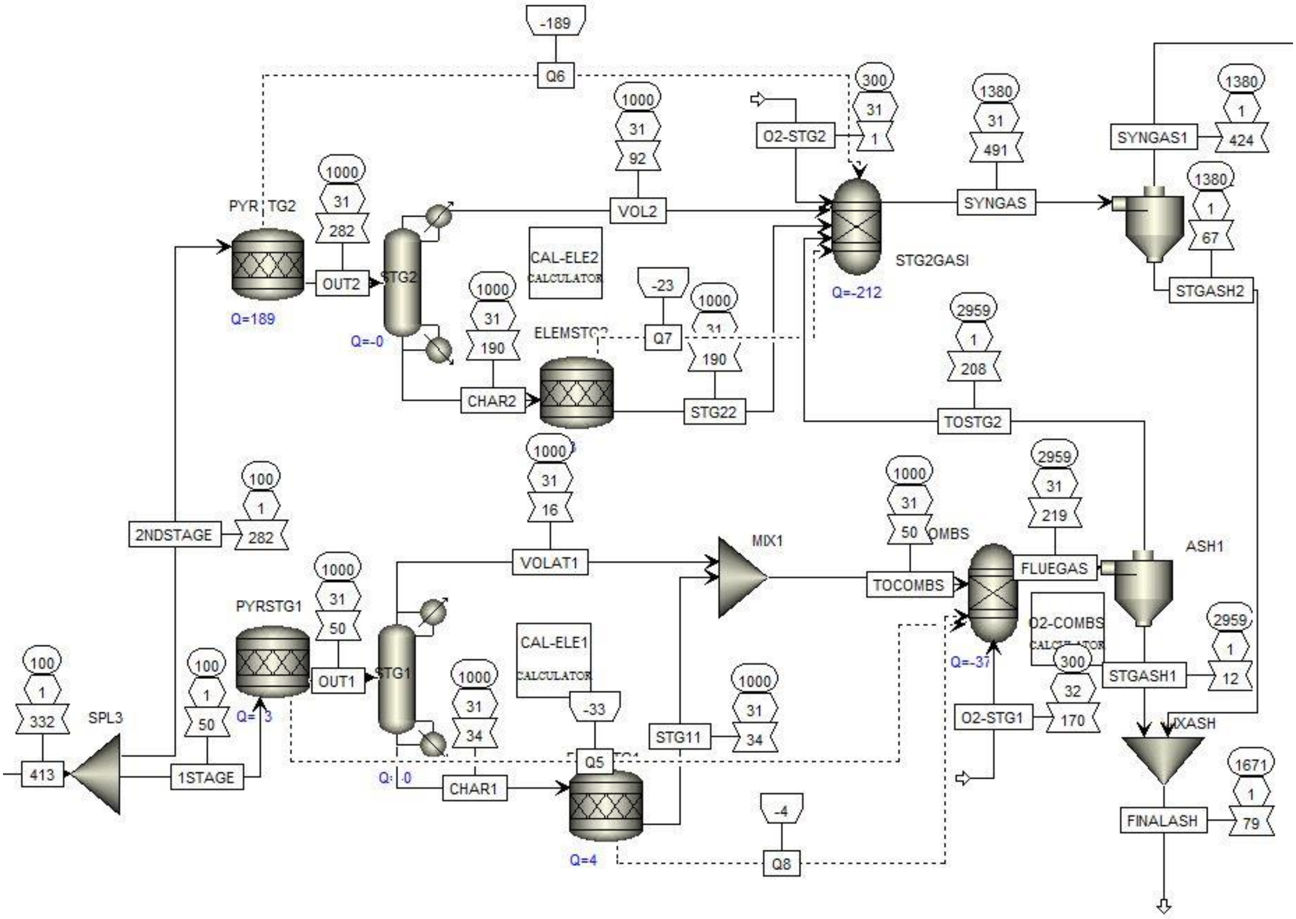
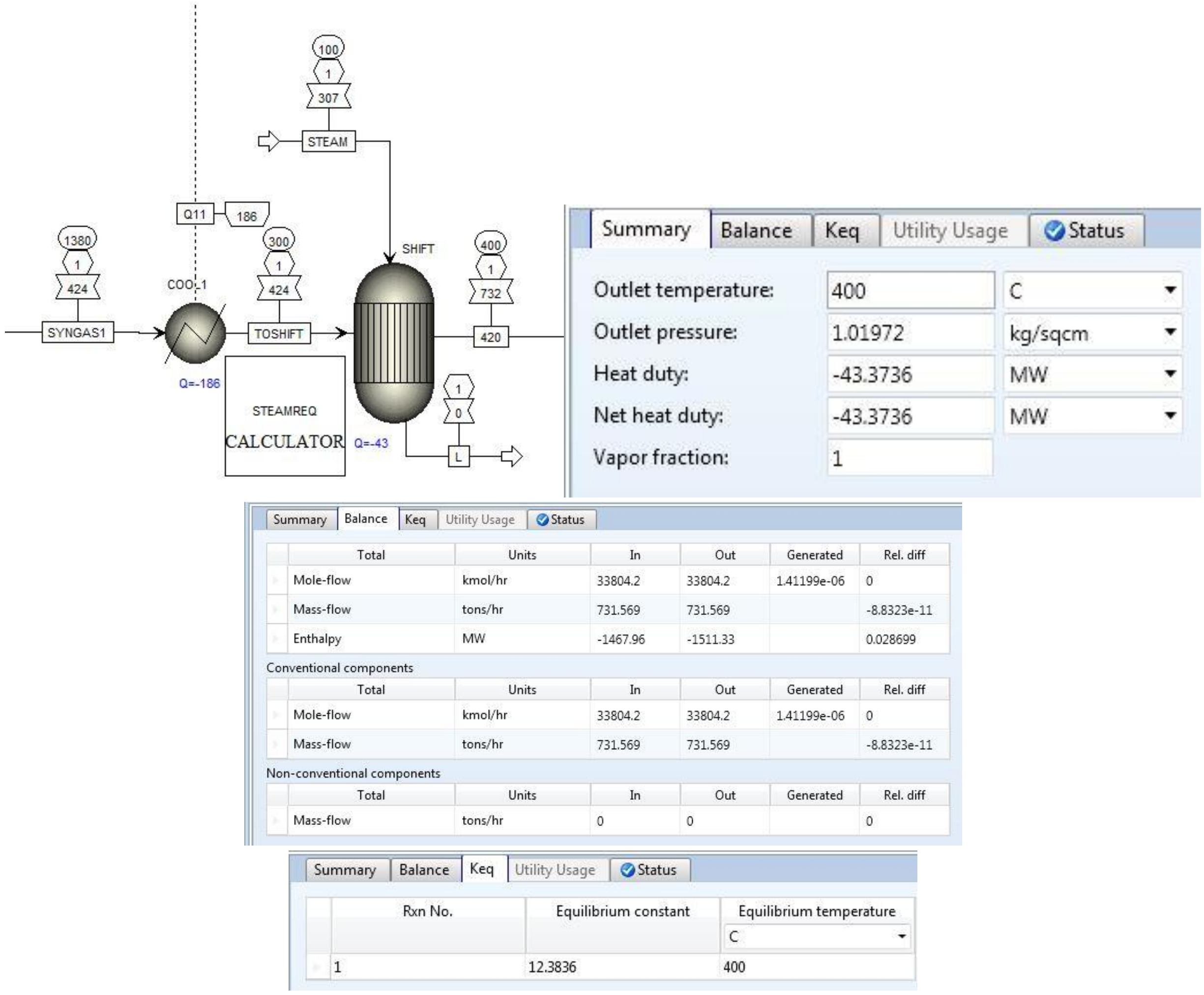


Fig 72: **D**etails of the two-stage Oxy-fired gasification unit

The flue gas from the combustor unit is carried to the second stage gasification zone. In this stage, Oxygen is injected to control the zone temperature close to 1400°C. 85% of the remaining dry coal is gasified in his zone. The Air separation unit (ASU) is not modelled in this flowsheet.

SHIFT reactor and heat recovery unit

The SHIFT reactor is modelled as a REquil reactor with specific 400°C reactor temperature. The steam requirement for this unit is calculated through a calculator block. A heater block is used to step down the syngas temperature to 300°C. The heat recovered from this block will be used for steam generation for the dryer block.



Outlet temperature: 400 C

Outlet pressure: 1.01972 kg/sqcm

Heat duty: -43.3736 MW

Net heat duty: -43.3736 MW

Vapor fraction: 1

Total	Units	In	Out	Generated	Rel. diff
Mole-flow	kmol/hr	33804.2	33804.2	1.41199e-06	0
Mass-flow	tons/hr	731.569	731.569		-8.8323e-11
Enthalpy	MW	-1467.96	-1511.33		0.028699

Conventional components

Total	Units	In	Out	Generated	Rel. diff
Mole-flow	kmol/hr	33804.2	33804.2	1.41199e-06	0
Mass-flow	tons/hr	731.569	731.569		-8.8323e-11

Non-conventional components

Total	Units	In	Out	Generated	Rel. diff
Mass-flow	tons/hr	0	0		0

Rxn No.	Equilibrium constant	Equilibrium temperature C
1	12.3836	400

Fig 73: Details of Shift reactor unit

Gas cleaning and CO_2 separation unit

The gas cleaning and CO_2 separation unit was modelled as a $CACO_3$ looping unit presented in **Fig 74**. An extensive study was carried out, and we choose this option. $CACO_3$ looping can meet dual purposes, the pollutant removal as well as the separation of CO_2 from the mainstream. The main advantage of this unit is it does not require fresh water stream as like Ammine scrubbing. There is two Gibbs reactors with two cyclone separator. In the carbonator reactor, the reaction enthalpy is exothermic and specified as 700°C. The stream of gases 420 enters the reactor and reacts with CAO.

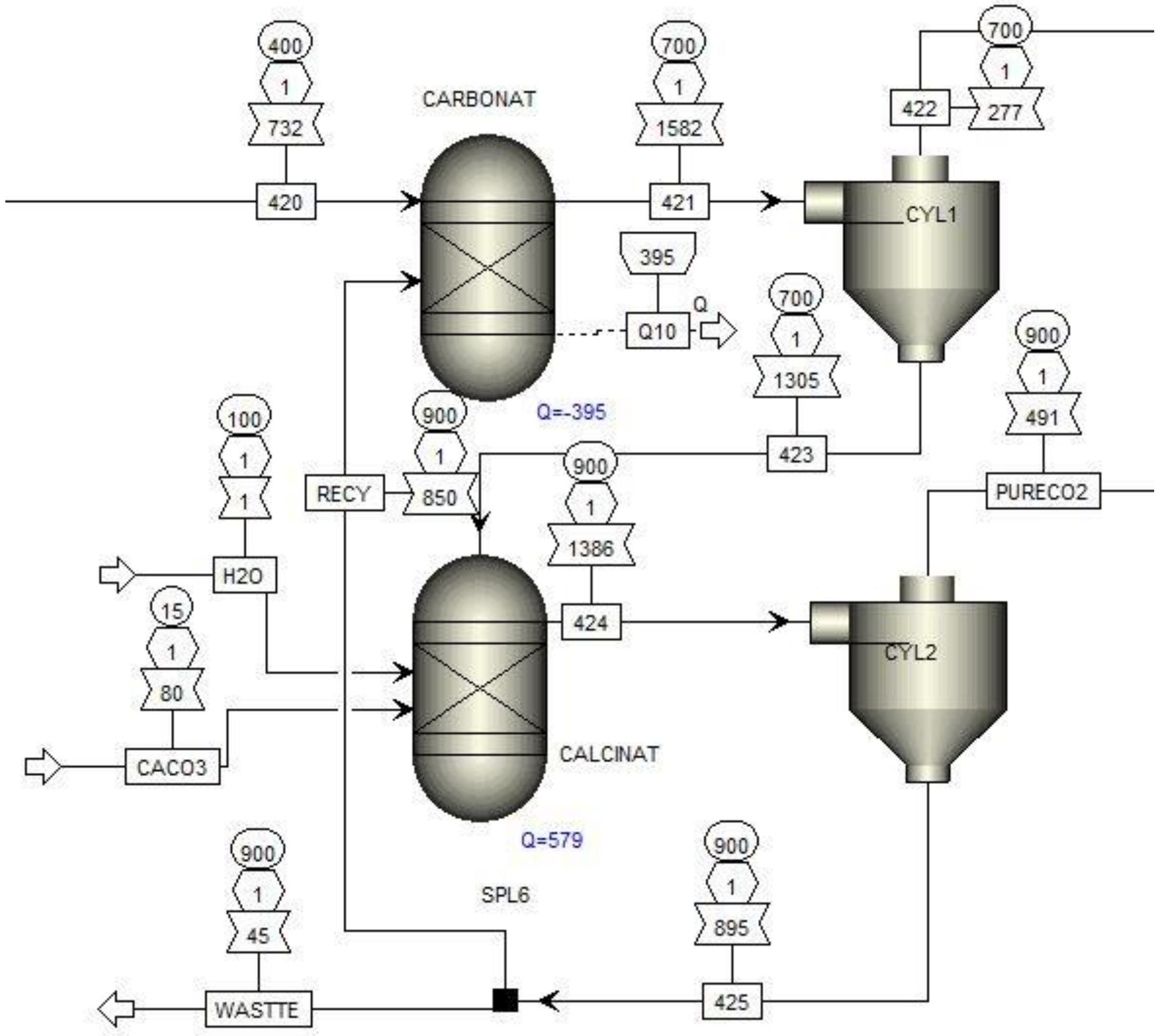


Fig 74: $CACO_3$ looping unit for gas cleaning and CO_2 separation

The reaction enthalpy in the CALCINATOR is endothermic. The heat requirement of this block is satisfied by the excess heat energy from the CARBONATOR and the excess energy from the combustor unit.

5% of the CAO was considered as waste. The entire unit was modelled in such a way so that in the recycle stream (USED CAO) there is no CAO going to CALCINATOR unit. A stream of steam (H2O) was added to the CALCINATOR so that it can drive the CO_2 and other pollutant gases from the system.

Heat recovery and PSA for H2 Separation

Energy is recovered from the clean gas and pure CO_2 stream through HEATER block as presented in **Fig 75**. The energy is then used for steam generation for the dryer. The H2 separation unit is considered as a component splitter.

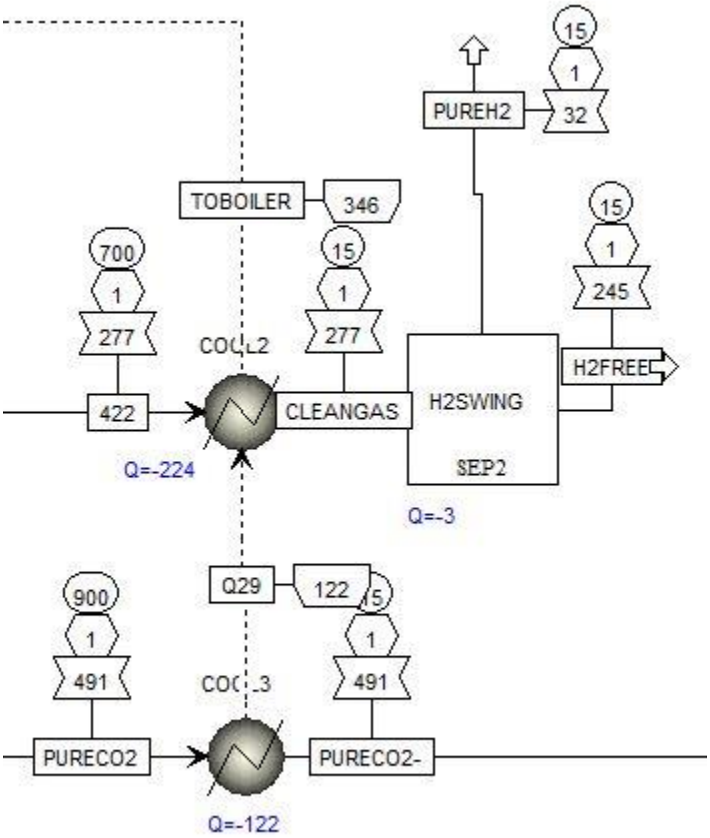


Fig 75: Heat Recovery and H2 Separation unit

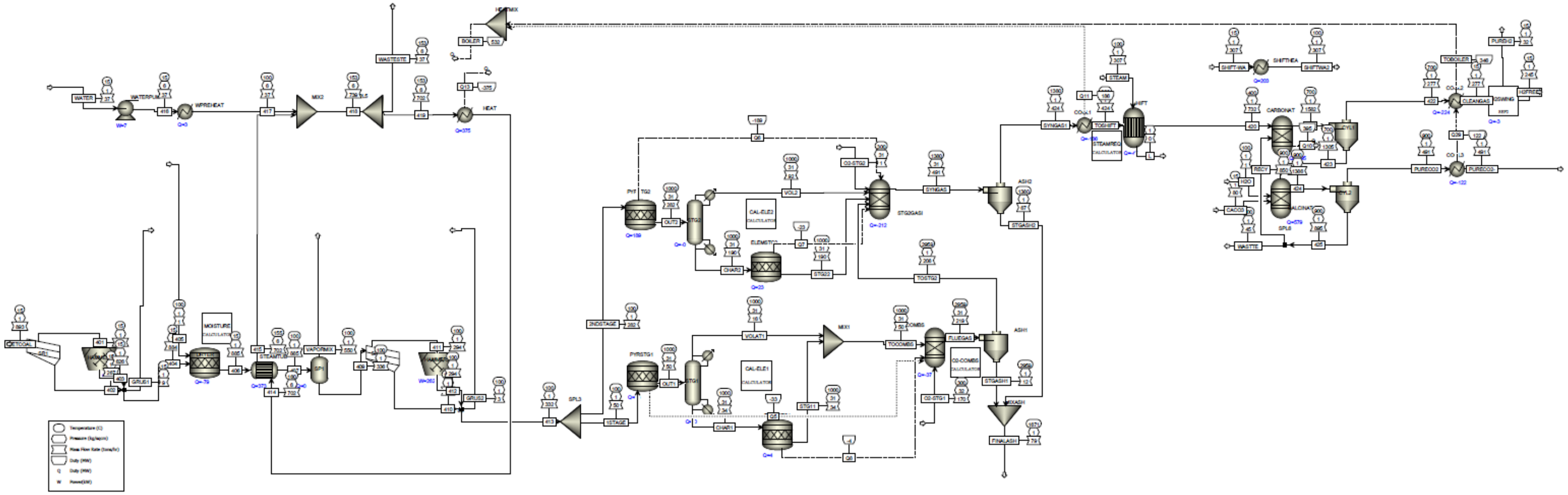
Table 6 present the stream results

Table 6

		WET COAL	STEAM	DRY COAL	O2-STG1	O2-STG2	CACO3	PURECO2	PUREH2
Mass Flow	TONS/HR	893	37	332.1645	169.5769	1	80	490.5215	31.99202
Mass Enthalpy	MW	-2386.01	-148.299	-239.391	11.19093	0.065993	-243.353	-1104.65	-1.14948
Mass flow rate									
H2	TONS/HR	0	0	0	0	0	0	0.000378	31.99202
CO	TONS/HR	0	0	0	0	0	0	1.321701	0
CO2	TONS/HR	0	0	0	0	0	0	487.1912	0
SO2	TONS/HR	0	0	0	0	0	0	1.011656	0
SO3	TONS/HR	0	0	0	0	0	0	8.50E-07	0
NO2	TONS/HR	0	0	0	0	0	0	0	0
NO	TONS/HR	0	0	0	0	0	0	0	0
S	TONS/HR	0	0	0	0	0	0	0	0
H2O	TONS/HR	0	37	6.79E-08	0	0	0	0.996609	0
C	TONS/HR	0	0	0	0	0	0	0	0
C2H6	TONS/HR	0	0	0	0	0	0	9.97E-32	0

CH4	TONS/HR	0	0	0	0	0	0	4.88E-16	0
H2S	TONS/HR	0	0	0	0	0	0	2.86E-05	0
C3H8	TONS/HR	0	0	0	0	0	0	5.36E-47	0
C3H6-2	TONS/HR	0	0	0	0	0	0	8.92E-41	0
O2	TONS/HR	0	0	1.85E-14	169.5769	1	0	1.62E-09	0
N2	TONS/HR	0	0	6.14E-14	0	0	0	0	0
NH3	TONS/HR	0	0	0	0	0	0	0	0
C10H8	TONS/HR	0	0	0	0	0	0	1.09E-100	0
C6H6O	TONS/HR	0	0	0	0	0	0	2.25E-63	0
TERT--01	TONS/HR	0	0	0	0	0	0	2.93E-86	0
C2H4	TONS/HR	0	0	0	0	0	0	2.63E-26	0
P-CRE-01	TONS/HR	0	0	0	0	0	0	1.04E-78	0
CAO	TONS/HR	0	0	0	0	0	0	0	0
CACO3	TONS/HR	0	0	0	0	0	80	0	0
CASO3	TONS/HR	0	0	0	0	0	0	0	0
CASO4	TONS/HR	0	0	0	0	0	0	0	0
CAS	TONS/HR	0	0	0	0	0	0	0	0
FE	TONS/HR	0	0	0	0	0	0	0	0
FE2O3	TONS/HR	0	0	0	0	0	0	0	0
FEO	TONS/HR	0	0	0	0	0	0	0	0
FE3O4	TONS/HR	0	0	0	0	0	0	0	0
COAL	TONS/HR	893	0	332.1645	0	0	0	0	0

CHAR	TONS/HR	0	0	0	0	0	0	0	0
ASH	TONS/HR	0	0	0	0	0	0	0	0



Appendix B Supplementary Report 2

The following included additional info from the Supplementary Report 2 on costing study by Bongers et al.

LCoP Calculation

There are various ways to calculate LCoP, and in this study we have used the following comprehensive formula taking into account financial assumptions (cost of capital, debt ratio, inflation rate, taxation life, and corporate/property taxes):

$$LCOP = \frac{\sum_{t=1}^{n} \frac{I_t + M_t + V_t + F_t}{(1+r)^t}}{\sum_{t=1}^{n} \frac{P_t}{(1+r)^t}},$$

where:

LCoP is the levelised cost of product

I_t is the investment expenditure in the year *t,* calculated with the WACC that takes into account tax depreciation using straight linear depreciation method.

M_t *is* the fixed O&M expenditure in the year (including sustaining capital) *t*

V_t is the variable O&M expenditure in the year *t*

F_t *is* the feedstock (fuel) expenditure in the year *t*

P_t is the product output in the year *t*

r is discount rate

n is the life of the plant

The capital cost of production capacities for various products, and O&M expenses are based on data from a number of public and non-public domain studies, which have been adjusted to appropriate plant designs and escalated to the Australian productivity factors as required.

IEA Hydrogen Activity List

The International Energy Agency Hydrogen Technology Collaboration Program (IEA-TCP) is a coordinated hydrogen research, development and demonstration program on a global basis. The have been focused on hydrogen research since the late 1970's. The completed tasks are listed in table below:

Completed Tasks		
Task 1	Thermochemical Production	1977-1988
Task 2	High Temperature Reactors	1977-1979
Task 3	Assessment of Potential Future Markets	1977-1980
Task 4	Electrolytic Production	1979-1988
Task 5	Solid Oxide Water Electrolysis	1979-1983
Task 6	Photocatalytic Water Electrolysis	1979-1988
Task 7	Storage, Conversion, and Safety	1983-1992
Task 8	Technical and Economic Assessment of Hydrogen	1986-1990
Task 9	Hydrogen Production	1988-1993
Task 10	Photoproduction of Hydrogen	1995-1998
Task 11	Integrated Systems	1995-1998
Task 12	Metal Hydrides for Hydrogen Storage	1995-2000
Task 13	Design and Optimization	1999-2001
Task 14	Photoelectrolytic Production	1999-2004
Task 15	Photobiological Production	1999-2004
Task 16	Hydrogen from Carbon-Containing Materials	2002-2005
Task 17	Solid and Liquid State Storage	2001-2006
Task 18	Integrated Systems Evaluation	2004-2006
Task 19	Hydrogen Safety	2004-2010
Task 20	Hydrogen From Waterphotolysis	2004-2007
Task 21	BioHydrogen	2005-2009
Task 21	BioInspired Hydrogen	2010-2014
Task 22	Fundamental and Applied Hydrogen Storage Materials Development	2006-2012
Task 23	Small Scale Reformers for OnSite Supply of Hydrogen (SSR for Hydrogen)	2006-2011
Task 24	Wind Energy and Hydrogen Integration	2006-2011
Task 25	High Temperature Hydrogen Production Processes	2007-2011
Task 26	WaterPhotolysis	2008-2011

Task 27	Near-Term Market Routes to Hydrogen by Co-Utilization of Biomass as a Renewable Energy Source with Fossil Fuels	2008-2011
Task 28	Large Scale Hydrogen Delivery Infrastructure	2010-2014
Task 29	Distributed and Community Hydrogen	2010-2014
Task 30	Global Hydrogen Systems Analysis	2010-2014
Task 31	Hydrogen Safety	2010-2013

MEDIA RELEASE
MINERALS COUNCIL OF AUSTRALIA

Using Australian coal to power hydrogen's future

Statement from Tania Constable, Chief Executive Officer

The Federal Opposition's National Hydrogen Strategy is a positive development for Australia's future energy policy.

Hydrogen offers an exciting new option as part of a diverse future global energy mix.

Only last month, the COAG Energy Council endorsed the National Hydrogen Roadmap and established the Hydrogen Working Group.

Australia and other major advanced economies such as Japan are already making large investments in this energy source.

Given Australia's significant coal resources, our country is well-placed to become a global producer of hydrogen in the future.

Converting Australian coal into hydrogen offers one of the best ways to provide a steady and reliable energy source that is not dependent on the weather.

A zero emission hydrogen solution could be delivered with advanced carbon capture and return technologies.

Australia is already leading the way. One of the world's largest hydrogen projects, now underway in the Latrobe Valley in Victoria, has attracted $500 million in investment, including from a consortium of leading Japanese companies along with the Victorian, Australian and Japanese governments.

This major project in Victoria will use Latrobe Valley brown coal and carbon capture and return technology and has the potential to make Victoria and Australia a leading global producer of hydrogen.

ends

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The costs in this document where finalised in February 2019.

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