Pre-Feasibility Study for Record Ridge Magnesia Production Plant

Prepared for:

West High Yield (WHY) Resources Ltd.

WEST HIGH YIELD
RESOURCES

ALBERTA, CANADA

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Do hereby certify that:

1. I reside at 4412 Rue Sainte-Emilie, Montreal, Québec H4C3L9.
2. I am a graduate from South Australian Institute of Technology with a B.Sc. Degree in Metallurgy (1986), a graduate of University of Toronto with a PhD. In Metallurgy (1993) and I have practiced for over 35 years.
3. I am a fellow of the “Australasian Institute of Mining & Metallurgy” (FAusIMM) (Membership Number 320930).
4. I am a Project Manager with Kingston Process Metallurgy Inc., which was incorporated in 2002.
5. I have not visited the property and the region in preparation of the report.
6. I have read the definition of "qualified person" set out in the National Instrument 43-101 (NI 43-101) and certify that as a result of my education, affiliation with a professional association (as defined in NI 43-101) and past relevant work experience, I fulfill the requirements to be a "qualified person" for the purposes of NI 43-101. I have been involved in metallurgical operations, engineering, construction and development, financial evaluation and senior management in the metals and minerals industry for over thirty five years.
7. I have no personal knowledge as of the date of this certificate of any material fact or change, which is not reflected in this report.
8. I am the author of Section 17 – Recovery Methods and Section 21 – Capital and Operating Costs and I have read the 2015 PEA report of SRK.
9. Neither I, nor any affiliated entity of mine, is at present, under an agreement, arrangement or understanding or expects to become, an insider, associate, affiliated entity or employee of West High Yield Resource or any associated or affiliated entities.
10. Neither I, nor any affiliated entity of mine own, directly or indirectly, nor expect to receive, any interest in the properties or securities of West High Yield Resource or any associated or affiliated companies.

Dated this 17th day of November, 2022.
(Signed and Sealed)

KWat

____
Kevin Watson, PhD., FAusIMM.

Florent Baril, P. Eng. Metallurgical Engineer

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FLORENT BARIL, ENG., B.Sc

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1. I reside at 624, Jean Deslauriers, Condo 17, Boucherville, Québec J4B 8P5.
2. I am a graduate from Laval University, Québec with B.Sc. Degree in Metallurgy (1954), and I have practiced for over 50 years.
3. I am a member of the "Ordre des Ingénieurs du Québec" (O.I.Q.) (Quebec Order of Engineers) (Membership Number 6972).
4. I am the Owner and President of Bumigeme Inc, a firm of consulting engineers, which has been incorporated in 1994.
5. I have not visited the property and the region in preparation of the report.
6. I have read the definition of "qualified person" set out in the National Instrument 43-101 (NI 43-101) and certify that as a result of my education, affiliation with a professional association (as defined in NI 43-101) and past relevant work experience, I fulfill the requirements to be a "qualified person" for the purposes of NI 43-101. I have been involved in mining operations, engineering, construction and development, financial evaluation and senior management in the mineral industry and engineering for over fifty years.
7. I have no personal knowledge as of the date of this certificate of any material fact or change, which is not reflected in this report.
8. I am the author of sections 22 and I have read the report of the Pre-Feasibility Study performed by KPM in collaboration the Tenova and KON Chemical Solutions e.U. in 2022 and the PEA of SRK in 2015.
9. Neither I, nor any affiliated entity of mine, is at present, under an agreement, arrangement or understanding or expects to become, an insider, associate, affiliated entity or employee of West High Yield Resource or any associated or affiliated entities.
10. Neither I, nor any affiliated entity of mine own, directly or indirectly, nor expect to receive, any interest in the properties or securities of West High Yield Resource or any associated or affiliated companies.

Dated this 9th day of November, 2022.
(Signed and Sealed)

Florent Baril, P. Eng.

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17.1 Introduction and Objectives

Experimental work conducted at Kingston Process Metallurgy Inc. (KPM) has demonstrated that a high purity (>99 wt%) magnesia (MgO) can be produced from the West High Yield Resources (WHY) Record Ridge deposit using a hydrochloric acid (HCl) leaching process1. In addition, promising results were obtained regarding production of a silica by-product from the leach residue. Based on the results of this work, a potential commercial plant flowsheet was developed, and a high-level mass and energy balance was calculated for a commercial plant.

It was recommended that WHY proceed with a two-stage commercialization pathway. In the first stage, a semi-commercial Demonstration plant would be designed, built and operated. The design would be customized using the results of the recent experimental work with the Record Ridge ore. In the second stage, a full-scale Commercial plant would be designed and built. To support this commercialization pathway, a prefeasibility study (PFS) for MgO production was conducted by KPM in collaboration with Tenova, Austria, and KON Chemical Solutions e.U. This report details the methodology and results of the PFS.

The overall objective of this work was to support commercialization of WHY's Record Ridge project. The specific objectives of the work were to prepare:

1. A detailed design and economic evaluation of the demonstration plant.

2. A high-level design and economic evaluation of the commercial plant.

17.2 Demonstration Plant

The Demonstration plant design is based off a plant designed by Tenova. The motivations for using this design are that it has been proven in operation and the plant capacity (~2,000 tpa product MgO) is of sufficient size to enable product qualification with potential buyers. Using this design would lower the project's technical and market risks, and engineering cost.

[1] Report- WHY - FSV - Stage 2 (P1804) KPM 211221

17.2.1Basic Design Data

17.2.1.1 Site Conditions

Table 1: Demonstration Plant Basic Design Data – Site Conditions.

Item	Unit	Value
Plant Location:		
Plant Location:	Location TBD in southern British Columbia	
Plant Elevation:	m amsl	1043
Temperature:		
Minimum:	°C	-15
Maximum:	°C	30
Atmosphere Pressure:		
Minimum:	mBar	1000
Maximum:	mBar	1030
Average:	mBar	1020
Humidity:		
Summer:	% rel.	44
Winter:	% rel.	85
Rainfall:		
Max. annually:	mm	1000
Max. monthly:	mm	120
Max. daily:	mm	50
Snow:		
Load:	kg/m^2	300
Wind:		
Predominant wind speed:	m/s	1.5
Max. Wind Speed	m/s	4.5
Earthquake Zone:		
Seismic zone		1

17.2.1.2 Raw Material

The chemical composition of the raw material feed[2] to the plant is shown in Table 2. The particle size distribution is assumed to be P80: 235-300µm.

Table 2: Demonstration Plant Basic Design Data – Raw Material.

Item	Weight %
SiO_2	44.9
MgO	39.8
FeO	10.2
Al_2O_3	0.6
Cr_2O_3	0.5
CaO	0.4
NiO	0.3
H_2O	3.3

17.2.1.3 Plant Operating Time

The plant operating time assumptions are shown in Table 3. These reflect the steady-state operating potential of the plant assuming that preventive maintenance is conducted according to the equipment Specifications.

Table 3: Demonstration Plant Basic Design Data – Plant Operating Time.

[2] Report- WHY - FSV - Stage 2 (P1804) KPM 211221

Item	Unit	Value
Total available time (24x365 days)	h	8,760
Production off-time	h	0
Yearly planned down time	h	490
Operating Time (net/year)	h	8,270
Estimated net availability	%	87
Net Production Time per year	h	7,200

17.2.2 Utilities and Consumables

17.2.2.1 Utilities (Owner's Scope)

Table 4: Demonstration Plant Basic Design Data – Utility Requirements.

Utility	Unit	Value
Natural Gas		
Net calorific value	MJ/Nm^3	>36
Pressure	bar	~ 2
Electrical energy		
LV power supply	VAC	3x460
UPS	VAC	120
Cycles	Hz	60
Deionized Water		
Pressure	bar	> 4
Conductivity	µS/cm	< 20
Industrial Water		
Pressure	bar	> 4
pH		> 7
Total hardness	mg/l $CaCO_3$	< 300
Quality		Free of impurities
Cooling Water		
Temperature in	°C	25
Temperature out	°C	35
Pressure	bar	> 3
Caustic Soda Solution (scrubber)		
Concentration	Wt%	Approx. 20
Caustic Thiosulfate Solution (scrubber)		
concentration	Wt%	~ 20
Compressed Air/Instrument Air		
Pressure	bar	> 6
Temperature	°C	< 40
Dew Point	°C	< -2
Quality		acc. ISO 8573-2/-1
Steam		
Pressure	bar	~ 6
Temperature	°C	~ 160
Chlorine Gas		
Purity	%	> 99
Fresh Acid		
HCl Content	%	> 30
Fluoride	ppm	< 5
Heavy Metal (Pb)	ppm	< 5

17.2.3Process Flow Diagram and Process Description

The Process Flow Diagrams (PFDs) for the Demonstration plant are shown in Figure 1 - Figure 4. General layout drawings of the Demonstration plant equipment and building are shown in Figure 5 and Figure 6. The key unit processes and underlying technical basis are described in the following sections.

17.2.3.1 Chemistry of the Leaching Process

In the leaching process, the raw material supplied to the system (Serpentine) is leached with a hydrochloric acid (HCl) solution. The raw material consists of the two main parts: silicon dioxide (SiO_2) and magnesium oxide (MgO) as well as different accompanying substances such as iron oxide (Fe_2O_3) and aluminum oxide (Al_2O_3). Furthermore, the serpentine also contains trace elements such as calcium oxide (CaO).

During leaching, the following chemical reactions take place and the metal oxides soluble in HCl are converted into the corresponding metal chlorides. In this process, water and heat are produced in addition to the chlorides.

$$MgO + 2HCl \rightarrow MgCl_2 + H_2O$$
$$Al_2O_3 + 6HCl \rightarrow 2AlCl_3 + 3H_2O$$
$$Fe_2O_3 + 6\ HCl \rightarrow 2\ FeCl_3 + 3H_2O$$
$$CaO + 2\ HCl \rightarrow CaCl_2 + H_2O$$

The SiO_2 has low solubility in HCl and therefore remains as a solid.

The leaching is conducted in two stages. In stage one, the Serpentine is leached with an excess of HCl and undissolved SiO_2 is separated from the leach solution by filtration. In stage two, additional Serpentine is added to consume the remaining HCl. The solution from stage two is transferred to the precipitation stage. Any unleached solid from stage 2 is recycled back to stage one.

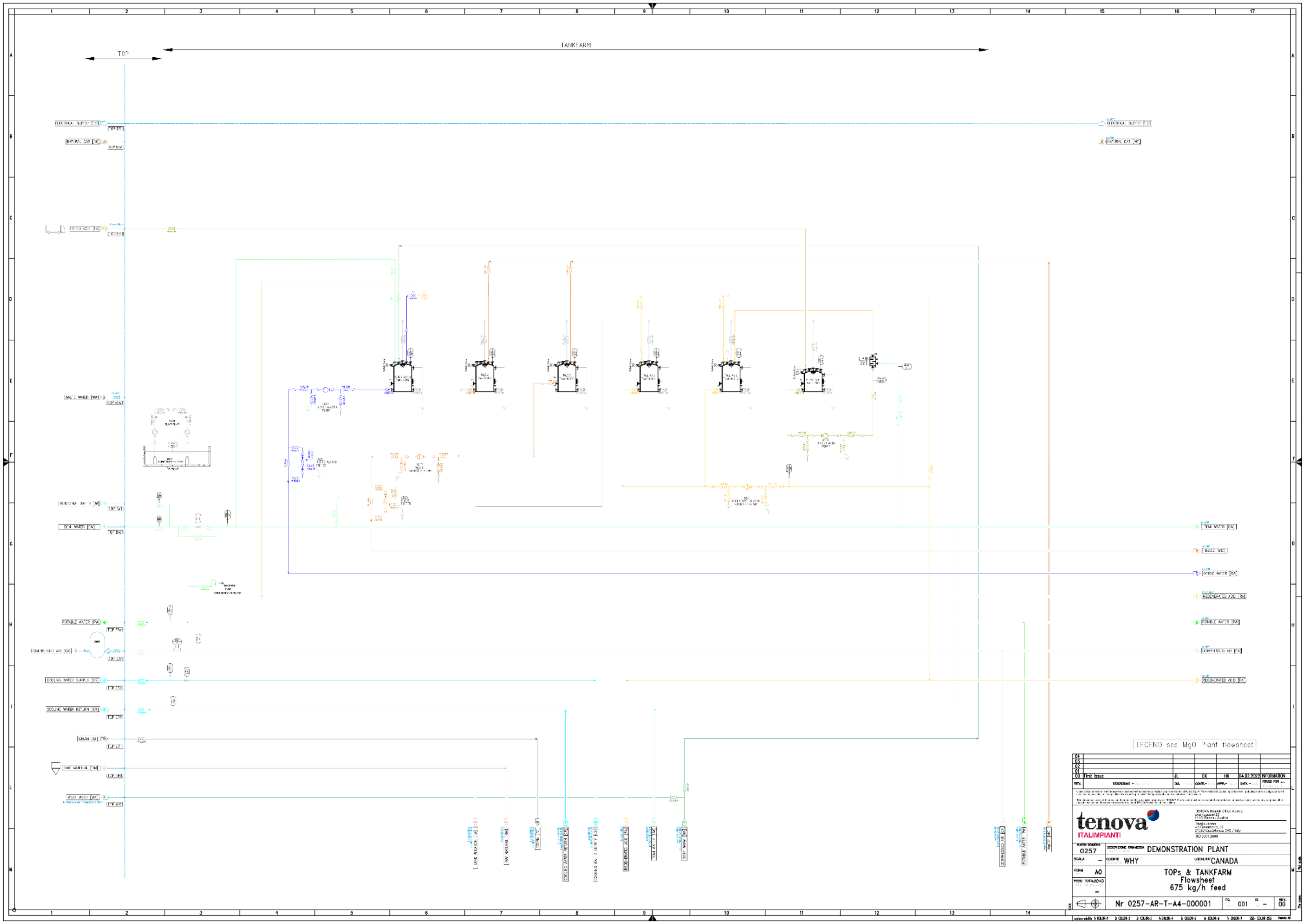
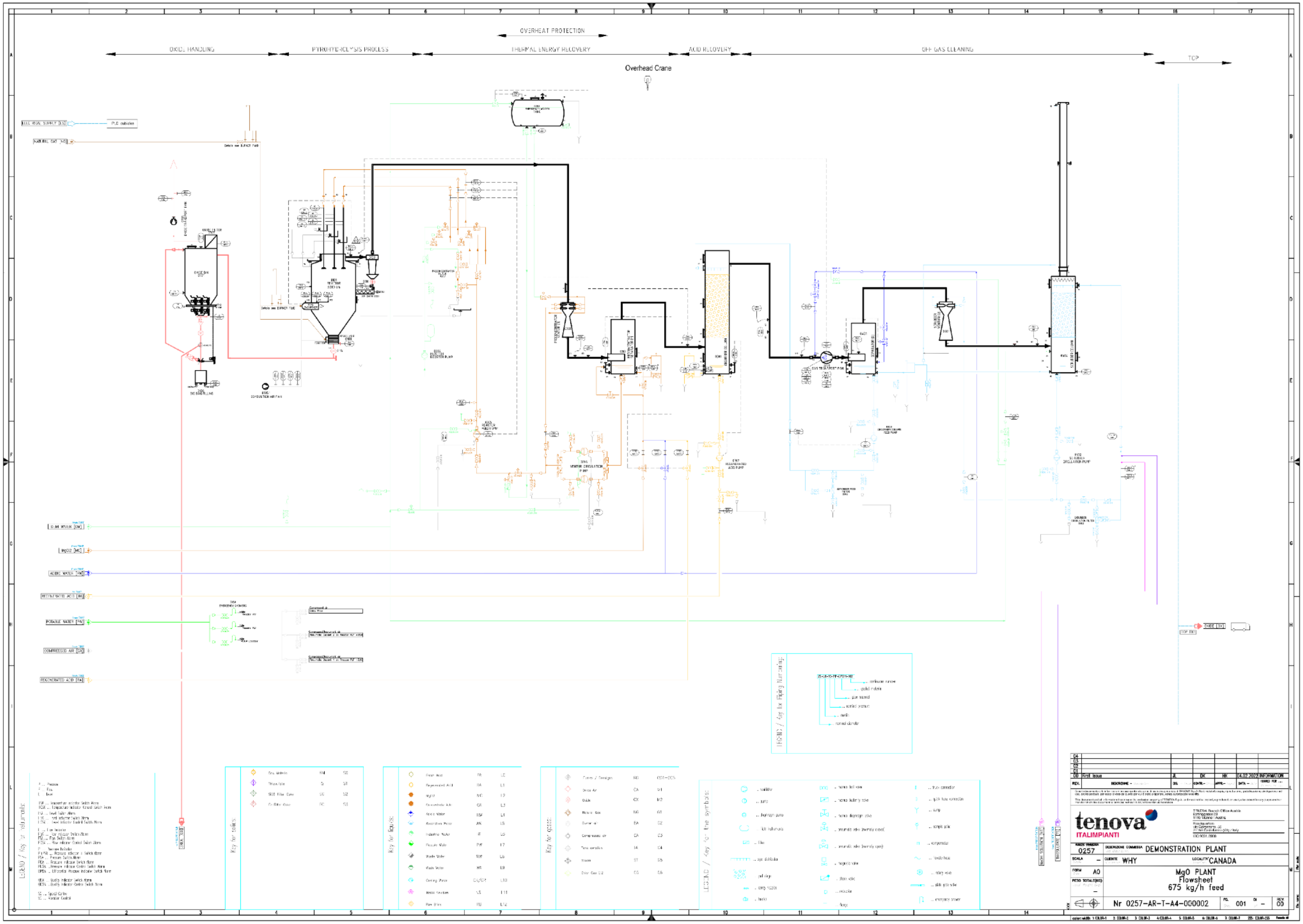


Figure 1: Process Flow Diagram for the Demonstration Plant – TOPs and Tank Farm.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PYROHYDROLYSIS PROCESS
THERMAL ENERGY RECOVERY
ACID RECOVERY
OFF GAS CLEANING
TCP
Overhead Crane
tenova
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DEMONSTRATION PLANT
WHY
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MgO PLANT
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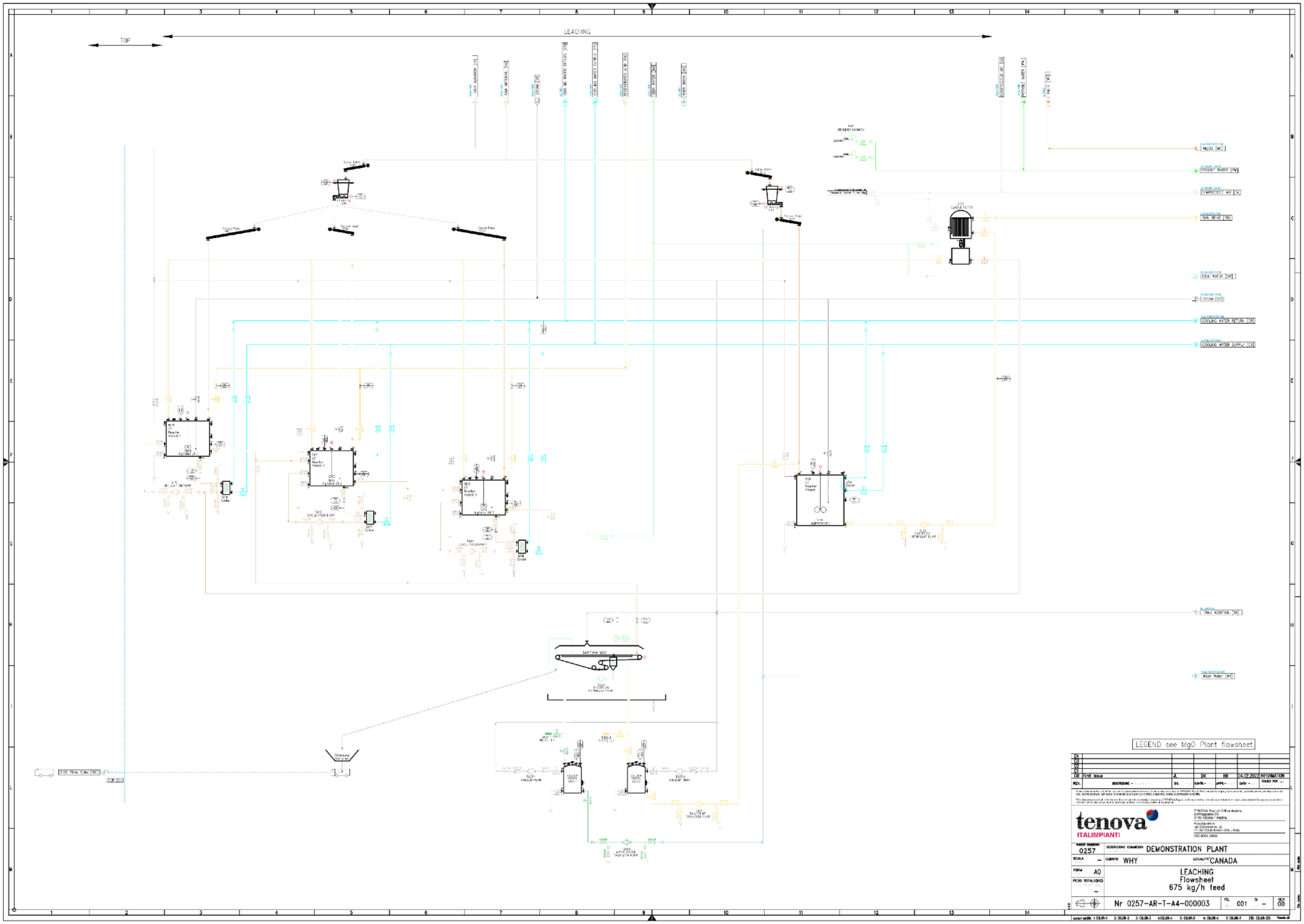
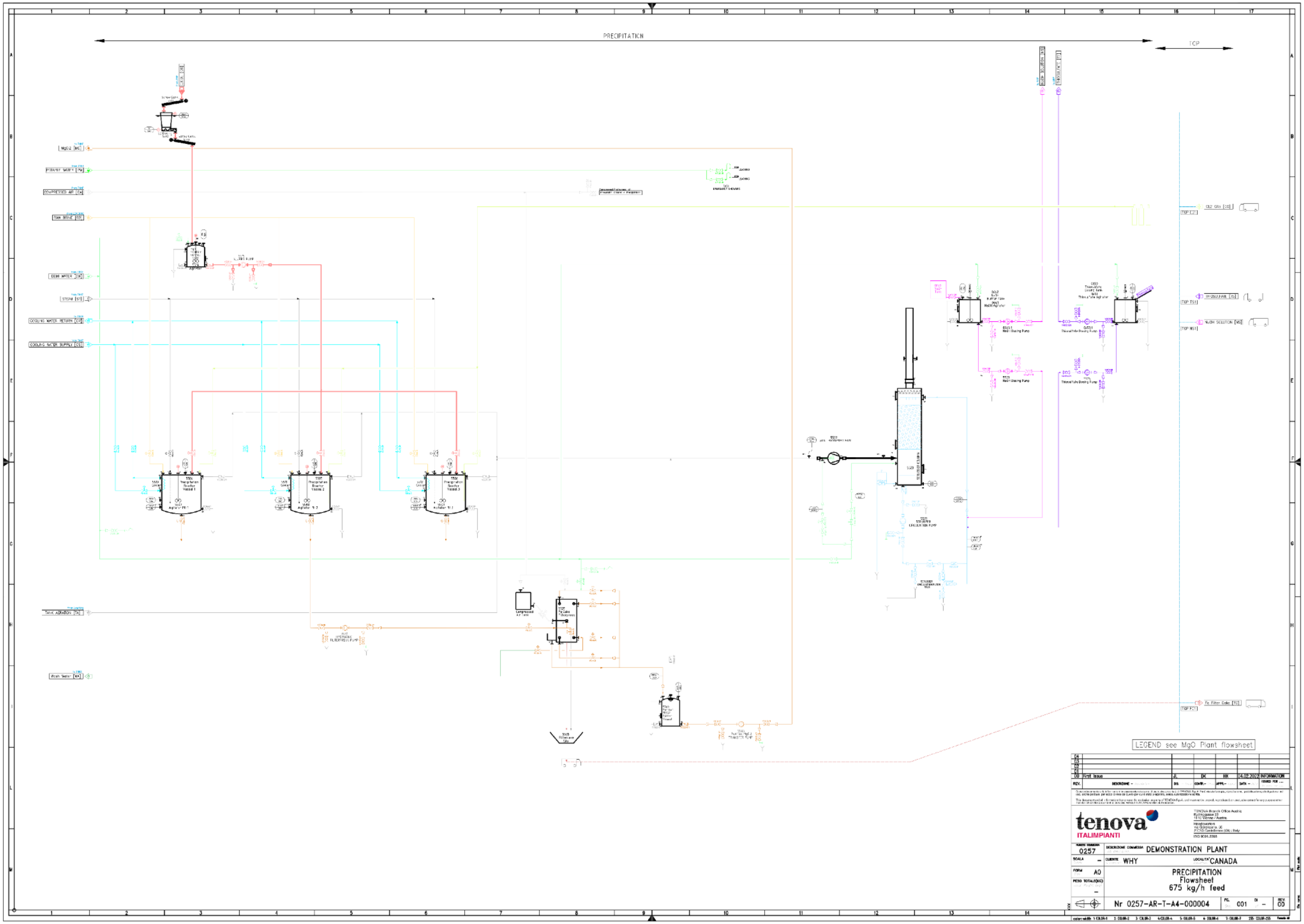


Figure 3: Process Flow Diagram for the Demonstration Plant – Leaching.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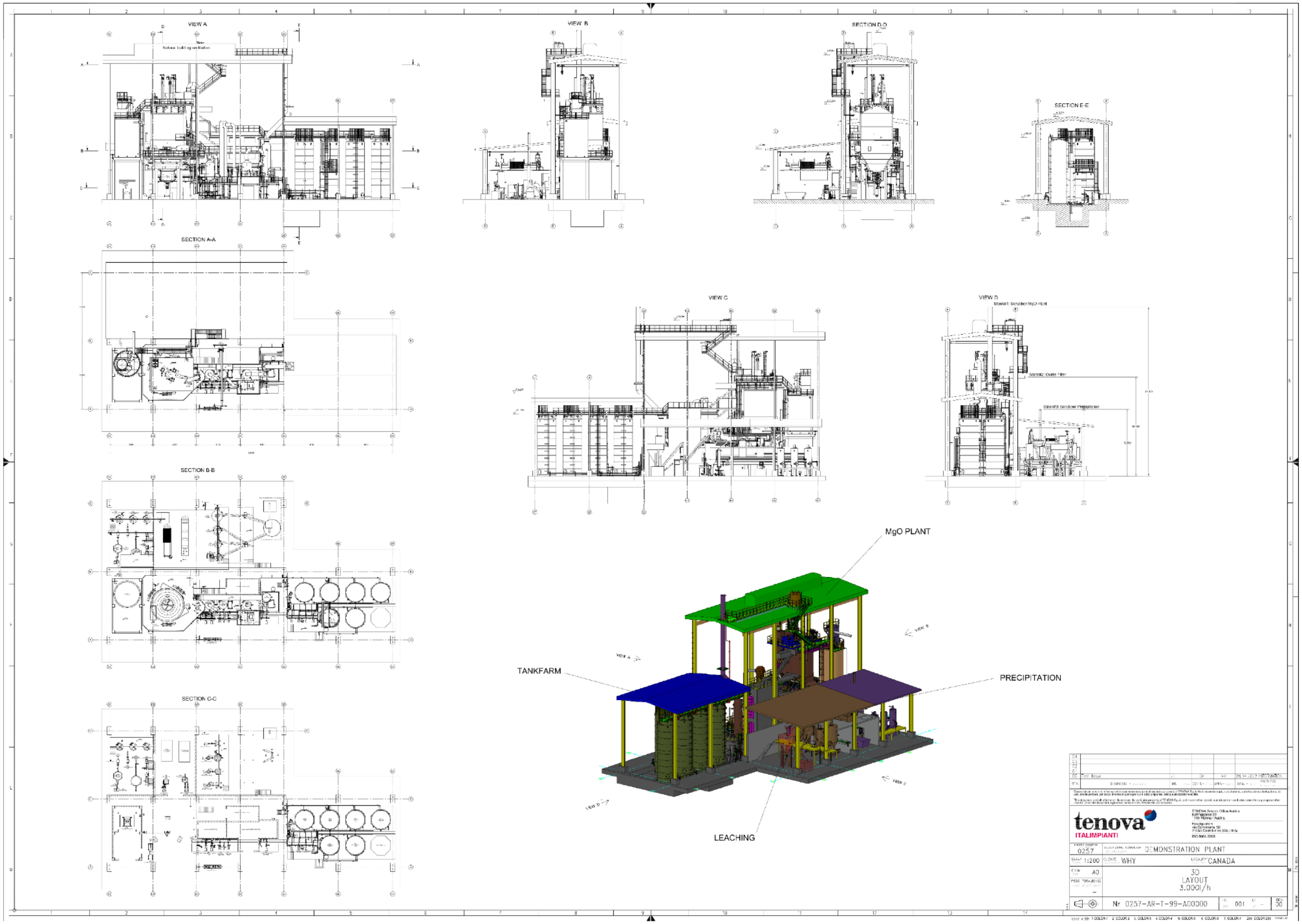


Figure 5: General Layout of Demonstration Plant – Equipment.

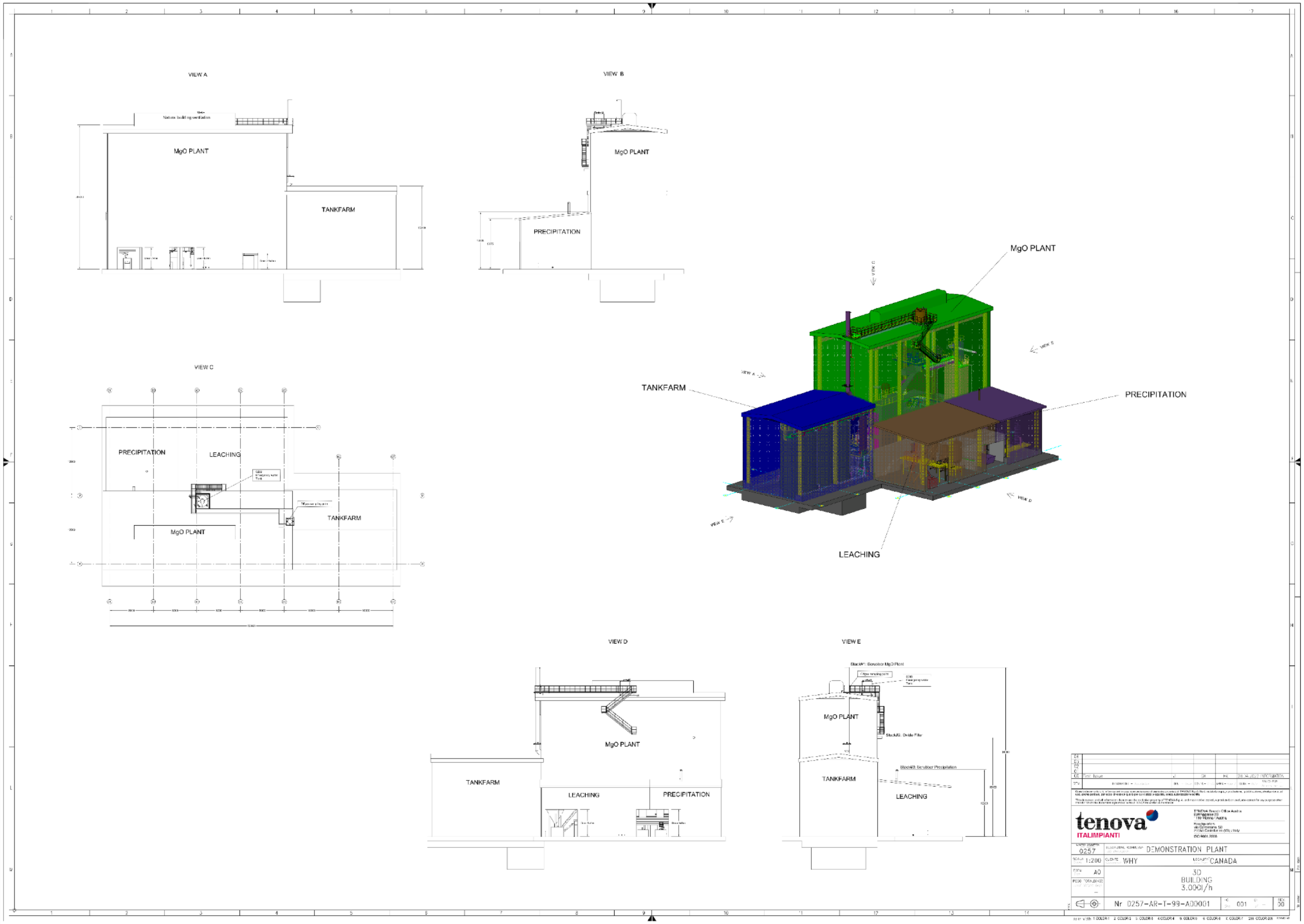


Figure 6: General Layout of Demonstration Plant – Building.

17.2.3.2 Chemistry of the Precipitation Process

During precipitation, the leach solution is treated with an oxidant (chlorine gas) to oxidize iron present as ferrous ions (Fe^{++}) as per:

$$2FeCl_2 + Cl_2 \rightarrow 2\ FeCl_3$$

MgO supplied from pyrohydrolysis is then added to increase the pH value of the solution leading to the precipitation of hydroxides and subsequent purification of the $MgCl_2$ solution as per:

$$2FeCl_3 + 3MgO \rightarrow 2\ Fe(OH)_3 + 3\ MgCl_2$$

$$2AlCl_3 + 3\ MgO \rightarrow 2\ Al(OH)_3 + 3\ MgCl_2$$

The precipitation process operates in batch mode.

17.2.3.3 Chemistry of the Pyrohydrolysis Process

In pyrohydrolysis, the purified $MgCl_2$ solution is sprayed as fine droplets into a reactor operated at 400 to 600°C. The water in the droplets is evaporated and decomposition of the $MgCl_2$ occurs according to the following reaction:

$$MgCl_2 + H_2O \rightarrow MgO + 2HCl$$

The resulting MgO falls onto the reactor bottom from where it is reclaimed still hot. The hot HCl gas is removed from the top of the reactor and pre-cleaned with a cyclone. The hot HCl gas is then cooled in a venturi using a $MgCl_2$ solution routed in a cycle. Water evaporates from the $MgCl_2$ solution and the $MgCl_2$ concentration increases.

Finally, the still gaseous HCl is absorbed by water in a counter-current scrubber resulting in a HCl acid solution with approximately azeotropic composition, termed "regenerated acid". The regenerated acid is then recycled back to the leaching process.

17.2.3.4 Description of the First Leaching Stage and SiO2 Separation

The leaching process will be implemented using three agitated vessels. To provide process flexibility and enable optimization trials to support process scale up for the Commercial plant, the vessels can be operated in batch or continuous operation mode. In both cases, the input battery limit is the serpentine feed as described in Table 2.

22.1.1.1.1 Batch Operation

In the respective vessels, the corresponding amount of regenerated acid from the pyrohydrolysis process step is supplied at the beginning of the cycle. Additionally, and if needed, a little HCl is added from tank storage for refreshing of the solution and replacement of acid loss (e.g. Cl^- loss as $CaCl_2$). The serpentine feed is then added to the vessel and the slurry is intensely mixed to suspend the solids, enhance liquid/solids contact and avoid dead zones. At the conclusion of leaching, the vessel is emptied and the acidic suspension consisting of SiO_2 solid and the generated metal chloride solution is transferred to solid/liquid separation by filtration. By having three vessels an alternating process can take place with the filter receiving material semi-continuously. The advantage of batch operation is that leaching can take place under tightly controlled conditions.

The leach vessels are connected to the central gas scrubber, which permits provide a pressure compensation during filling and emptying and ensures that no hazardous substances can escape untreated. In addition, each vessel is connected to a cooling water system to remove the heat from the exothermic leaching reactions and control the desired leaching temperature.

17.2.3.5 Continuous Operation

Continuous operation is achieved using a cascade approach. All the feed serpentine and the HCl acid is supplied to Vessel 1, which is connected to Vessel 2 by an overflow with a subsequent overflow connection from Vessel 2 to Vessel 3. The overflow from Vessel 3 is the final output from the first stage of leaching leach and it is connected to the filter. Each vessel is agitated and cooled as described above. The cascade approach results in an average total retention time of solid particles in the system. Three stages are used to reduce the probability of particles short-circuiting the system directly to the overflow and having insufficient leaching retention time.

The advantage of a continuous system is that no down-time is lost to filling and emptying the vessels, which increases overall system efficiency.

With both modes of leaching operation, a significant amount of free HCl remains in the leaching solution to ensure that highest possible extraction rate of solvable components.

The solid/liquid separation takes place by a washed belt filter with a movable vacuum trough that permits virtually continuous operation. The vacuum required for filtration is produced by a water ring vacuum pump. The vacuum leads to better separation of the liquid from the solid aiding filter cake drying. However, it leads to the undesired addition of ambient air to the system that is subsequently removed by vacuum vessels next to the filter. The filter is connected to the central scrubber to avoid emissions into the working environment.

The washed and still moist SiO_2 filter cake is emptied into in a filter cake container. The filtrate solution is removed using vacuum-resistant pumps that are designed for low suction-side pressures.

17.2.3.6 Description of the Second Leaching Stage

The solution from the first stage of leaching is transferred to a larger agitated leaching vessel in the second stage of leaching. Additional raw Serpentine is added, which consumes the remaining free acid. The vessel is agitated, cooled and connected to the central scrubbing system as described above with the first leaching stage vessels. To have a reasonable leaching reaction speed it is planned to apply an excess of serpentine raw material amount, which will not be fully leached.

The sold/liquid separation is executed in a candle filter and the solid fraction is transferred to the first leaching stage to complete the extraction process. The liquid fraction is transferred to the oxidation and precipitation stage.

17.2.3.7 Description of the Precipitation Stage

There are three agitated and temperature-controlled process vessels available for oxidation and precipitation of the filtrate from the second leaching stage. The cyclic operation of the three units permits quasi continuous operation. Each vessel is supplied with Cl_2 gas for the oxidation process and dosing of solid MgO for pH-controlled precipitation. Once the precipitation process has taken place, the $MgCl_2$ suspension is transported to a washed filter-press for solid-liquid separation. The purified and filtered $MgCl_2$ solution is transferred to buffer storage in the tank farm ahead of the roasting stage.

17.2.3.8 Description of the Roasting Stage

The first part of the roasting stage is a preheater/preconcentrator. The pure $MgCl_2$ solution is pumped from buffer storage in the tank farm to the venturi, where it is mixed with hot process gas up to 500°C. This causes evaporation of water contained in the solution, cooling of the process gas and concentrating the $MgCl_2$ solution. Cooling is essential so that rubberized process equipment or equipment made of PP or FRP can be used in the following process steps. Evaporation of water and concentration of the $MgCl_2$ solution increases the energy efficiency of the process.

The reactor feed pump transports the concentrated $MgCl_2$ solution to the spray reactor, where it is finely dispersed in the hot reactor space through four supply points. To avoid clogging of the spray nozzles, the reactor supply system is equipped with several fine filters. For the start and shut-down processes of the reactor, there is the option of spraying water into the reactor space. This water is provided by the reactor booster pump.

In the reactor, the pyrohydrolysis process, in which the $MgCl_2$ solution is decomposed to MgO and HCl, takes place. The gas containing hot HCl leaves the reactor at the furnace head and the MgO powder falls to the bottom of the reactor. The reactor bottom includes a finalizer that allows a controlled final temperature treatment. The MgO exits the roaster via a rotary valve in the finalizer and is pneumatically conveyed to the product bin. The product will be a fine-grained technical grade MgO with a purity >98% MgO.

The energy required for the pyrohydrolysis process is provided through natural gas burners that are placed tangentially at the reactor jacket. The natural gas is provided at line pressure and the combustion air is transported to the burners through a combustion air blower.

The HCl-containing process gas discharged at the top of the roaster contains a certain dust load that is separated in a cyclone. The collected dust is returned to the reactor space takes place via a rotary valve and a screw feeder, which ensures that no backflow impairs the efficiency of the cyclone and guaranteed materials flow.

After dust removal, the hot process gas flows through the preheater/preconcentrator. It is then transported to a counterflow absorber for capture and regeneration of HCl acid. The absorber is charged with acidic water, which absorbs the gaseous HCl to form a near azeotropic acid, which is stored in a buffer tank in the tank farm. The spray roaster and HCl absorber system operate at a slight negative pressure to ensure no fugitive emissions of HCl-containing gas.

After HCl absorption system, the process gas is cleaned of traces of chlorine and HCl using a wet alkaline scrubber. The effluent from the scrubber column is fed to the waste water treatment facility.

17.2.4Plant Mass and Energy Balance

The mass and energy balance for the Demonstration plant is summarized in Figure 7. The utility and reagent consumption values are summarized in Table 5 and the predicted process performance in Table 6.

Table 5: Demonstration Plant Mass and Energy Balance – Consumption and Connected Loads.

Utility	Consumption Value	Connected Value (Owner's Scope)
Natural gas	260 Nm^3/h	500 Nm^3/h
Electrical power	215 kWh/h	400 kVA
Chlorine gas	32 kg/h	-
Deionized water	500 l/h	1 m^3/h
Potable water	discontinously	5 m^3/h
Absorption water	2.900 l/h	5 m^3/h
Wash water	750 l/h	1.5 m^3/h
Sodium-thiosulphate	8 l/h	-
Caustic soda	5 l/h	-
Industrial water	discontinously	10 m^3/h
Compressed air	100 Nm^3/h	200 Nm^3/h
Cooling water dT=10°C	50 m^3/h	100 m^3/h

Table 6: Demonstration Plant Mass and Energy Balance – Operating Values.

Item	Unit	Value
Serpentine Feed		
Quantity	kg/h	500
Serpentine Acid Capture		
Quantity	kg/h	177
Chlorine Gas Consumption		
Quantity	kg/h	32
MgO recycle purification		
Quantity	kg/h	64
Regenerated Acid		
Efficiency *	%	101
Concentration	% (wt)	18
Mg^{++}	g/l	<5
Oxide		
Quantity (design load reactor)	kg/h	314
Sellable Quantity (product= reactor - recycle)	kg/h	250
MgO **	%	>96.5
Cl^- ***	%	<1.2
Exhaust Gas		
HCl	mg/Nm^3 dry	<20
MgO – dust (SPM)	mg/Nm^3 dry	<20
Cl_2	mg/Nm^3 dry	<5

*Some Cl^- is lost via wash water solution and chemically bound to the non-roastable chlorides. It is estimated that these losses will be made up, with a slight excess, by the use of Cl_2 gas for the oxidation step.
**It is anticipated that the main residual impurity will be Ca.
***It is anticipated that $CaCl_2$ will be the main source of residual chlorides.

17.3 Commercial Plant

WHY's objective is to install and operate a plant treating 250,000 t/y ore from the Record ridge deposit. Assuming an on-time of 7,200 h/y, this equates to a plant capacity of 34.7 t/h ore feed. The plant would consist of three main unit operations: 1. Leaching; 2. Precipitation; and 3. Spray-Roasting. The first two are stirred tank operations and the tank size can be readily scaled to the throughput capacity required. By contrast, there is a limit on the maximum size for an individual $MgCl_2$ spray roaster that results in a maximum throughput capacity of approximately 25,000 l/h $MgCl_2$, which for the Record Ridge feed is equivalent to a maximum feed capacity of 6.5 t/h or 46,800 t/y ore. Therefore, a 250,000 t/y ore capacity plant would require five spray roasters.
For prefeasibility study, it was assumed that the commercial plant would be scaled up progressively based on installation of 6.5 t/h ore processing modules as illustrated in Figure 8. This strategy minimizes the initial capital outlay and would enable a progressive market entrance
The TEA for the Commercial plant presented below is based on installing a greenfield plant of capacity of 6.5 t/h ore or 46,800 t/y ore.

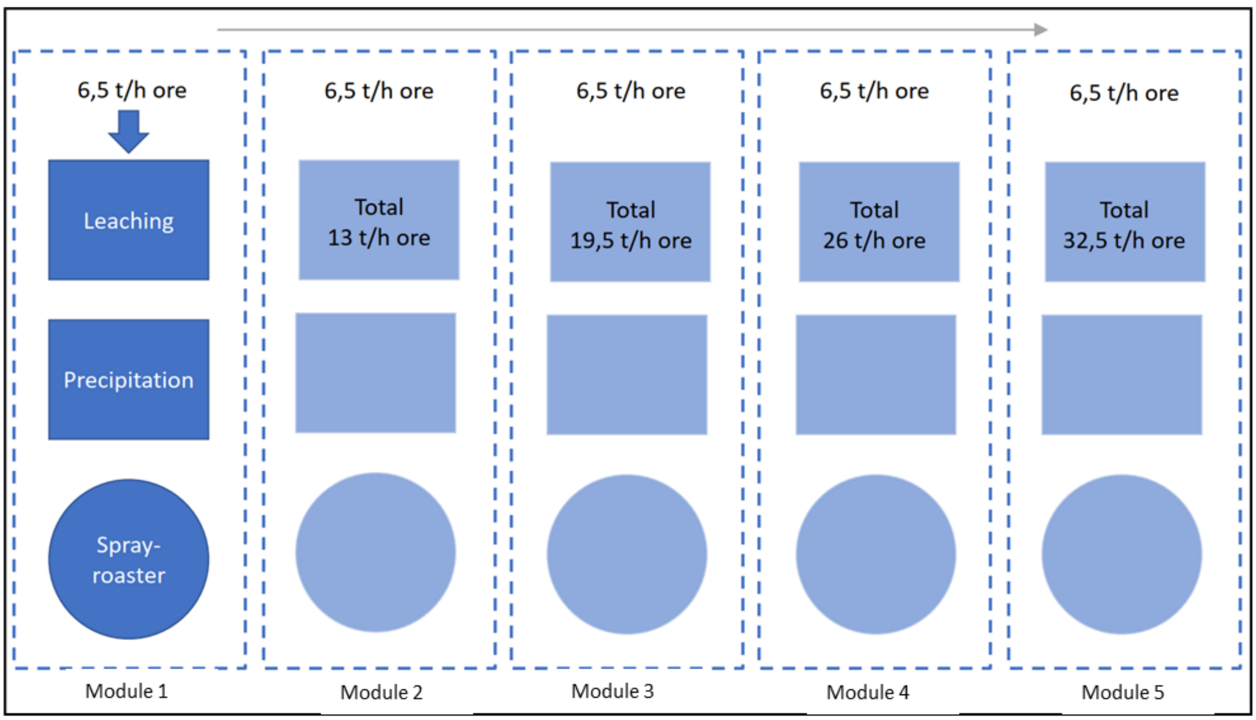


Figure 8: Baseline Commercial Plant Implementation Strategy.

Given the scalability of the Leaching and Precipitation unit operations, it would be possible to vary the implementation strategy depending on the availability of financing and the market demand. An example of an alternative strategy is shown in Figure 9. In this alternative, an initial 6.5 t/h plant is installed followed by two subsequent expansions of 13 t/h each, resulting in a final 34.7 t/h plant capacity. These and other scenarios can be further investigated as the project proceeds and as WHY's market entrance strategy is developed.

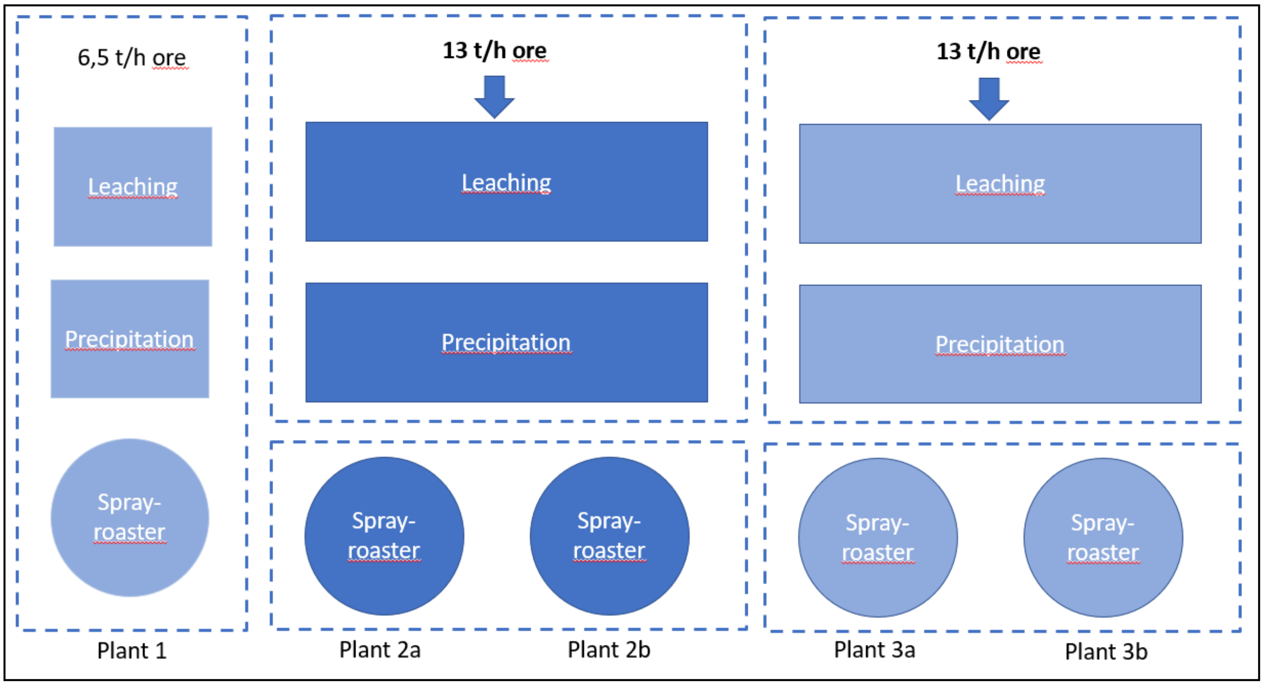


Figure 9: Alternative Commercial Plant Implementation Strategy.

17.3.1Basic Design Data

17.3.1.1 Site Conditions

It is assumed that the commercial plant will be in a location TBD in southern British Columbia. The pertinent site condition assumptions are given in Table 1.

17.3.1.2 Raw Material

It is assumed that the commercial plant will receive the same raw material feed as the demonstration plant and as detailed in Table 2.

17.3.1.3 Plant Operating Time

The commercial plant operating time assumptions are shown in Table 3. These reflect the steady-state operating potential of the plant assuming that preventive maintenance is conducted according to the equipment Specifications.

17.3.1.4 Utilities and Consumables

17.3.2Utilities (Owner's Scope)

Table 7: Commercial Plant Basic Design Data – Utility Requirements.

Utility	Unit	Value
Natural Gas		
Net calorific value	MJ/Nm^3	>36
Pressure	bar	~2
Electrical energy		
LV power supply	VAC	3x460
UPS	VAC	120
Cycles	Hz	60
Deionized Water		
Pressure	bar	>4
Conductivity	µS/cm	<20
Industrial Water		
Pressure	bar	>4
pH		>7
Total hardness	mg/l $CaCO_3$	<300
Quality		
Cooling Water		
Temperature in	°C	25
Temperature out	°C	35
Pressure	bar	>3
Caustic Soda Solution		
Concentration	Wt%	20
Caustic Thiosulfate Solution		
concentration	Wt%	20
Compressed Air/Instrument Air		
Pressure	bar	>6
Temperature	°C	<40
Dew Point	°C	<-2
Quality		ISO 8573-2/-1
Steam		
Pressure	bar	~6
Temperature	°C	~160
Chlorine Gas		

Purity	%	>99
Fresh Acid		
HCl Content	%	>30
Fluoride	ppm	5
Heavy Metal (Pb)	ppm	5

17.3.3 Process Flow Diagram and Process Description

The Process Flow Diagrams (PFDs) for the Commercial plant are shown in Figure 11 - Figure 14. General layout drawings of the Commercial Plant equipment and building are shown in Figure 15, Figure 16 and Appendix 2. The key unit operations and underlying technical basis will be the same as for the Demonstration plant and as described in Section 3.3.

17.3.4 Plant Mass and Energy Balance

The mass and energy balance for the Commercial plant is summarized in Figure 17. The utility and reagent consumption values are summarized in Table 8 and the predicted process performance in Table 9.

Table 8: Commercial Plant Mass and Energy Balance – Consumption and Connected Loads.

Utility	Consumption Value	Connected Value (Owner's Scope)
Natural gas	2,500 Nm^3/h	3,500 Nm^3/h
Electrical power	2,000 kWh/h	3 MVA
Chlorine gas	310 kg/h	-
Deionized water	2 m^3/h	5 m^3/h
Potable water	discontinously	5 m^3/h
Absorption water	29 m^3/h	50 m^3/h
Wash water	7.5 m^3/h	15 m^3/h
Sodium-thiosulphate	80 l/h	-h
Caustic soda	50 l/h	-
Industrial water	discontinously	10 m^3/h
Compressed air	300 Nm^3/h	500 Nm^3/h
Cooling water dT=10°C	500 m^3/h	750 m^3/h

Table 9: Commercial Plant Mass and Energy Balance – Operating Values.

Item	Unit	Value
Serpentine Feed		
Quantity	kg/h	4,800
Serpentine Acid Capture		
Quantity	kg/h	1,700
Chlorine Gas Consumption		
Quantity	kg/h	300
MgO recycle purification		
Quantity	kg/h	620
Regenerated Acid		
Efficiency*	%	101
Concentration	% (wt)	18
Mg^{++}	g/l	<5
MgO Product		
Quantity (design load reactor)	kg/h	3,000
Sellable Quantity (product= reactor - recycle)	kg/h	2,400
MgO**	%	>96.5
Cl^{-}***	%	<1.2
Exhaust Gas HCl	mg/Nm³ dry	<20
MgO – dust (SPM)	mg/Nm³ dry	<20
Cl_2	mg/Nm³ dry	<5

*Some Cl^- is lost via wash water solution and chemically bound to the non-roastable chlorides. It is estimated that these losses will be made up, with a slight excess, by the use of Cl_2 gas for the oxidation step. Optimizing the Cl balance will one of the objectives of demonstration plant.

**It is anticipated that the main residual impurity will be Ca.

***It is anticipated that $CaCl_2$ will be the main source of residual chlorides.

17.3.5 Product Specification

The product from the Commercial plant will be a fine-grained (98%) MgO. An illustrative datasheet for this MgO product is shown in Figure 10.

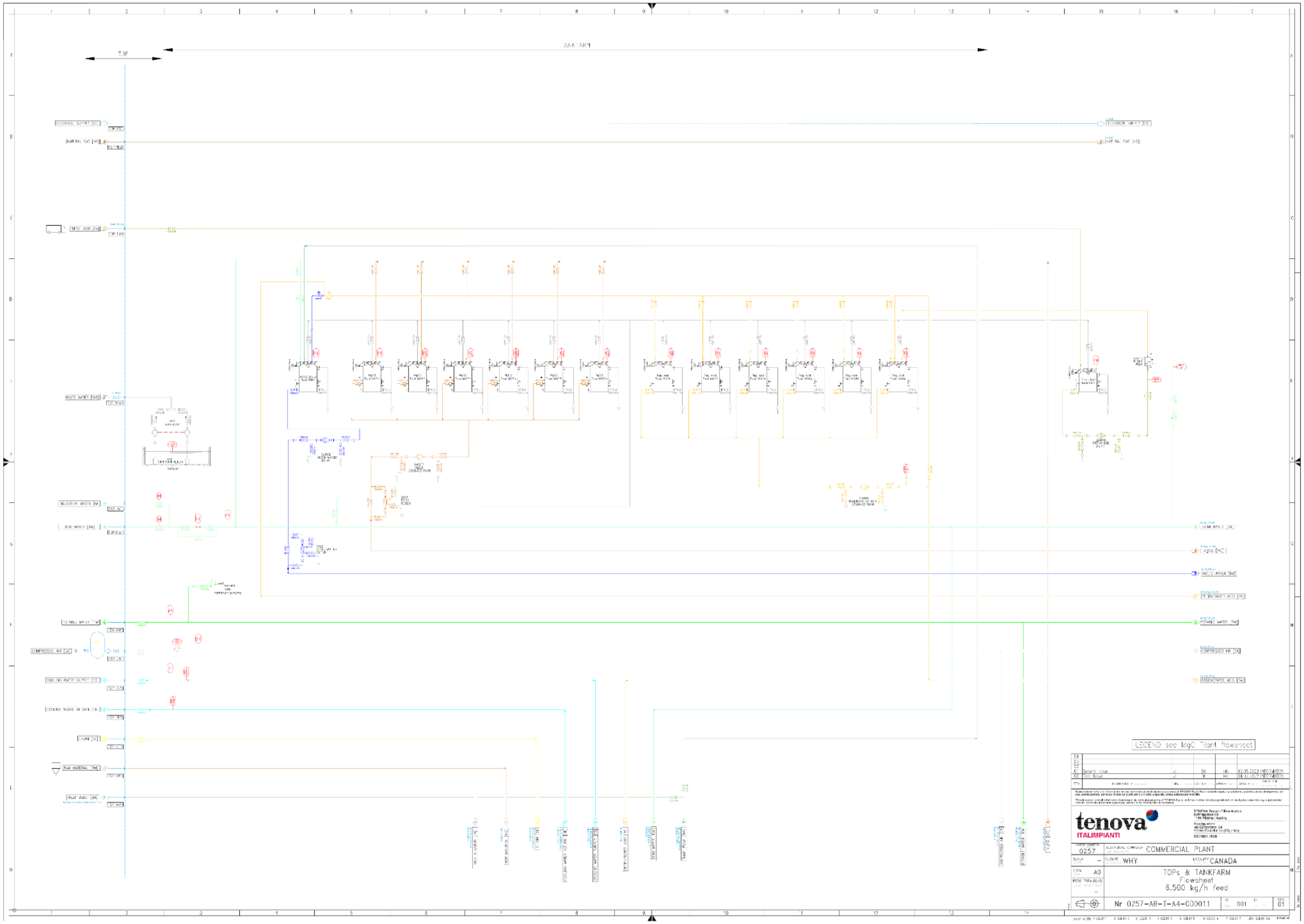
Magnesium Oxide (MgO) Powder

General Description	A fine (D50 ~ 2 micron) powder
Applications	A neutralization reagent in the chemical industry. A precursor material for production of high purity magnesium compounds such as magnesium hydroxide. A raw material for the pharmaceutical industry.
Packaging	The material is hydroscopic and is packed in 25 kg polyethylene bags and palletized for shipping.

Chemical Composition			
Item	**Unit**	**Typical**	**Specification**
MgO	%	97.5	>97.0
Cl^-	%	1.2	<1.5
CaO	%	0.45	<0.60
SO_4	ppm	300	<1000
Na_2O	ppm	300	<550
SiO_2	ppm	60	<200
NiO	ppm	50	<100
K_2O	ppm	30	<100
Fe_2O_3	ppm	30	<100
Al_2O_3	ppm	20	<50
Cr_2O_3	ppm	10	<30
MnO	ppm	10	<30

Chemical Properties			
Item	**Unit**	**Typical**	**Specification**
Moisture content (LOI)	%	1.5	<2.0
Specific surface area (BET)	m^2/g	4-5	

Figure 10: MgO Product Datasheet.


TOP
TANKFARM
LEGEND see MgO Plant flowsheet
01 Second Issue
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tenova
ITALIMPIANTI
0257
COMMERCIAL PLANT
WHY
CANADA
A0
TOPs & TANKFARM
Flowsheet
6.500 kg/h feed
Nr 0257-AR-T-A4-000011
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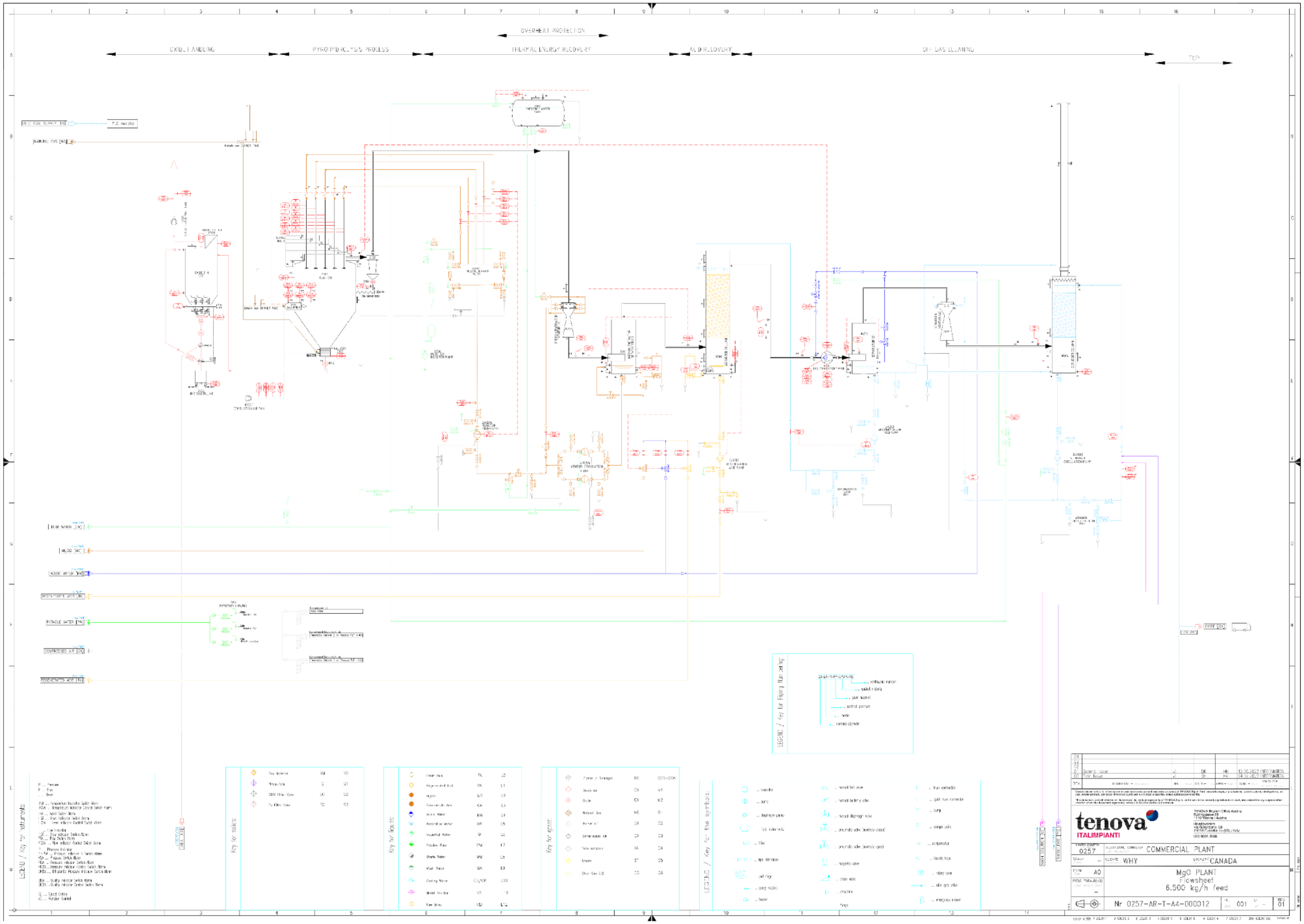


Figure 12: Process Flow Diagram for the Commercial Plant – MgO Plant.

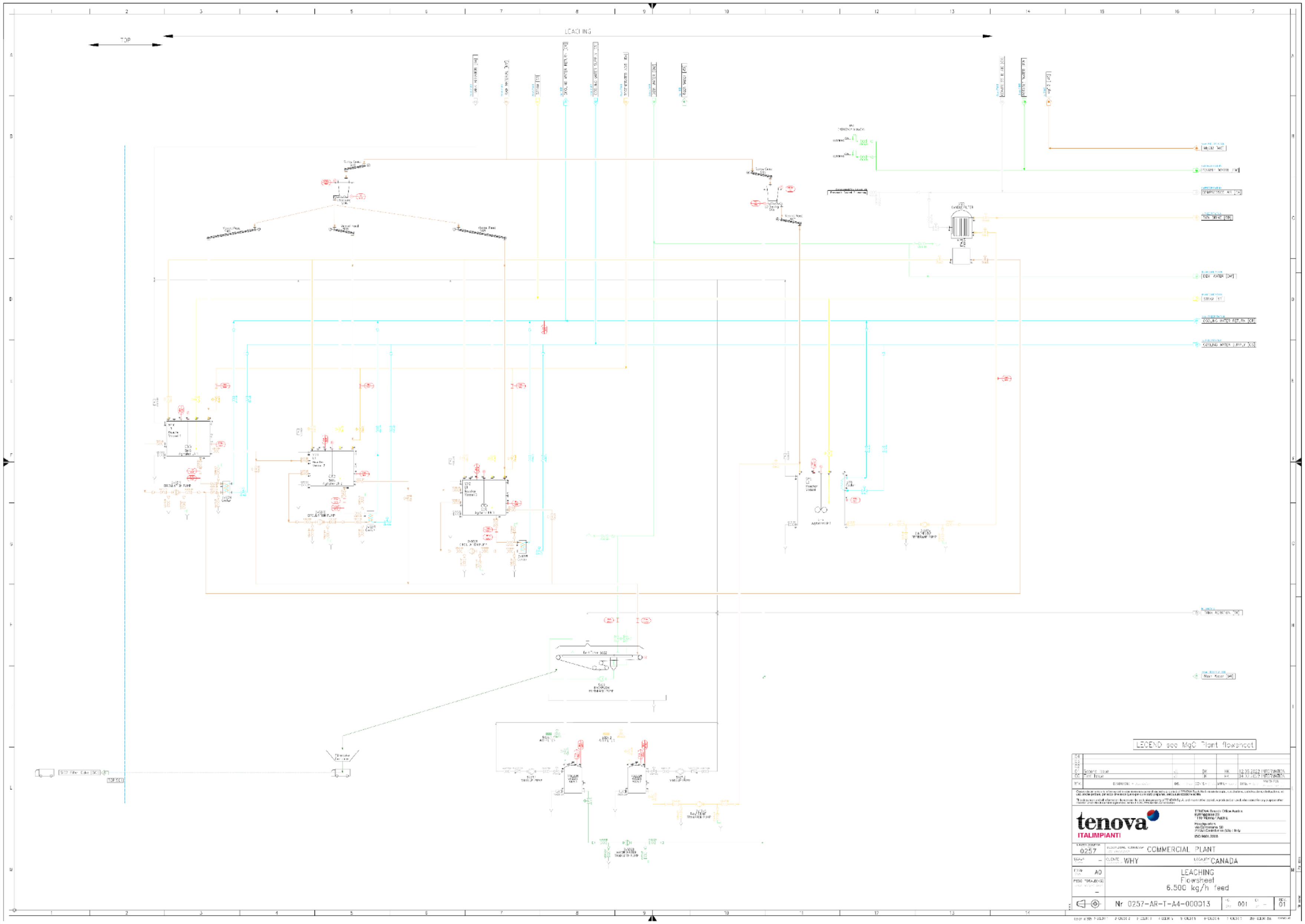
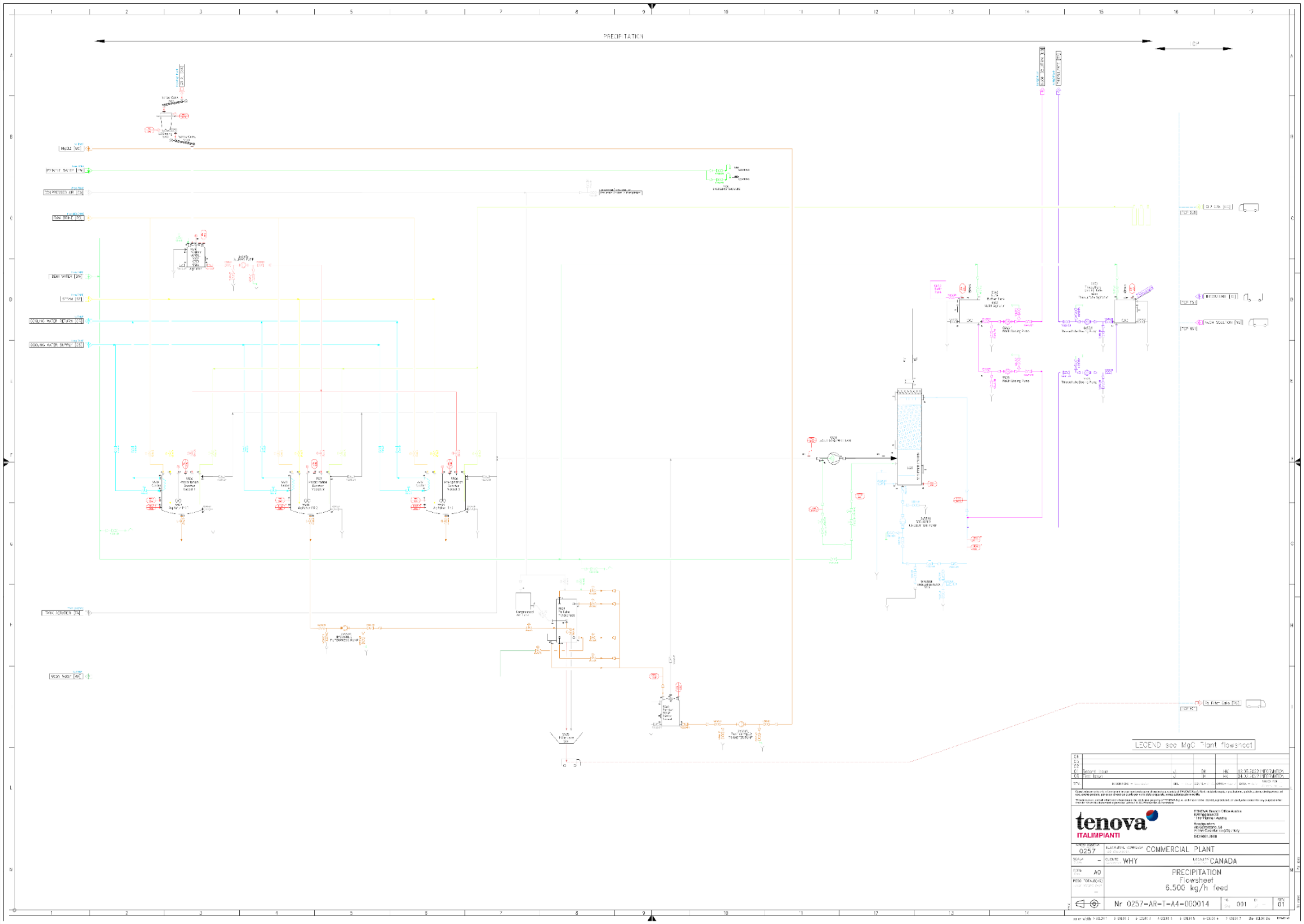


Figure 13: Process Flow Diagram for the Commercial Plant – Leaching.


PRECIPITATION
LEGEND see MgO Plant flowsheet
01 Second Issue
00 First Issue
tenova
ITALIMPIANTI
0257
COMMERCIAL PLANT
WHY
CANADA
A0
PRECIPITATION
Flowsheet
6.500 kg/h feed
Nr 0257-AR-T-A4-000014
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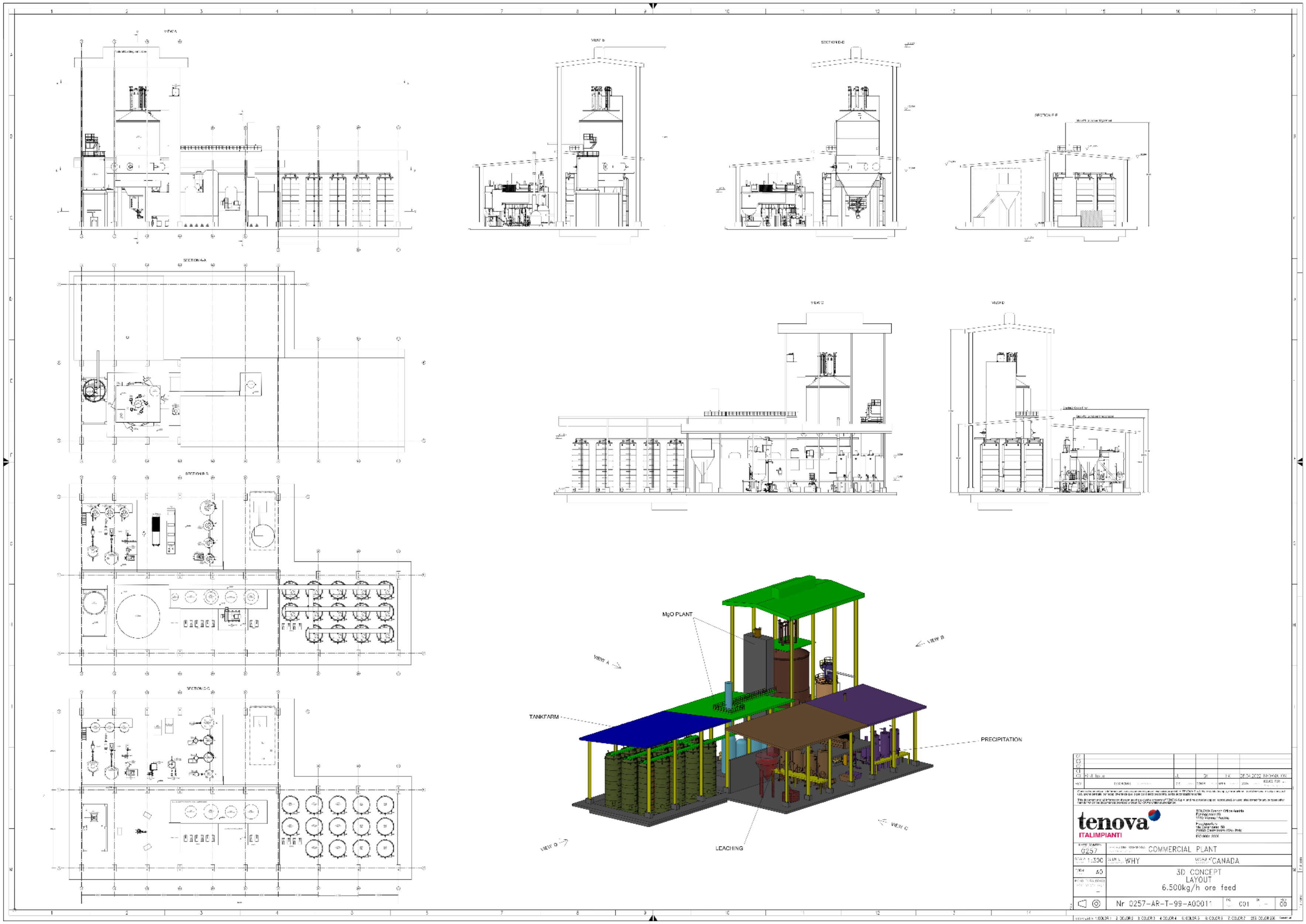


Figure 15: General Layout of Commercial Plant – Equipment.

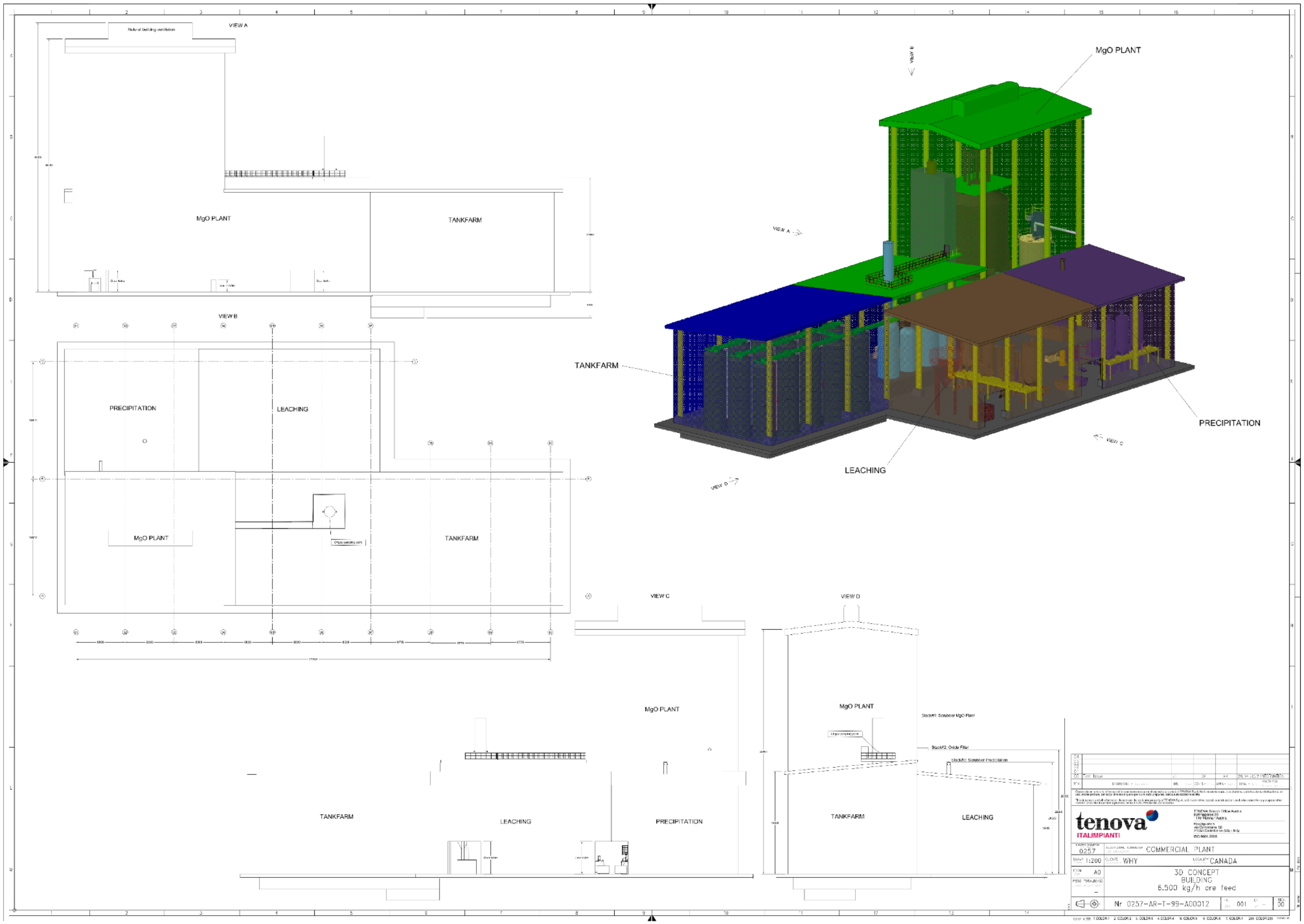


Figure 16: General Layout of Commercial Plant – Building.

21 CAPITAL AND OPERATING COST ESTIMATE

21.1 Capital Cost Estimate – Demonstration Plant

The equipment for the process were sized based on the output from the mass and energy model. An equipment list is given in Appendix 1. This equipment was then costed using quotes from suppliers and in-house data from similar recent projects. The equipment costs were then used to develop factored costs for installation, civils, piping, electrical and controls/instrumentation to develop a total direct capital cost. Indirect costs and a 15% contingency were added to the direct costs to estimate the total installed capital cost. Excluded from the capital cost estimate are:

- Costs for regulatory approvals.
- Environmental / geotechnical investigations and clean-up operations.
- Owner's costs.
- Spare parts.

The capital cost estimate for the Demonstration plant is summarized in Table 10. The total installed capital cost is estimated to be $CAD 27.9 million (-15/+20%).

Table 10: Capital Cost Estimate for the Demonstration Plant.

Major Units	**$CAD**
Leaching	3,271,000
Precipitation	2,292,000
Pyrohydrolysis	5,618,000
Tank farm	2,125,000
Balance of plant	487,000
Buildings	3,537,000
Total Direct Capital Cost	17,330,000
Indirect Costs	
EPCM & start-up services	5,226,000
Freight	1,103,000
Field indirect & first fill	565,000
Total Indirect Capital Cost	6,894,000
Total Direct and Indirect Costs	24,220,000
Contingency (15%)	3,640,000
Total Installed Capital Cost	**27,860,000**

21.2 Operating Cost Estimate – Demonstration Plant

The unit cost assumptions for the operating cost estimate for the Demonstration plant are summarized in Table 11. It is noted that the high prices for the reagents (i.e. sodium hydroxide, sodium thiosulfate and chlorine) reflect their relatively small consumption in the Demonstration plant. The unit costs for the same reagents for the Commercial plant, which will consume greater quantities, will be lower as indicated in Table 15.

Table 11: Unit Cost Assumptions for Demonstration Plant.

Item	Unit	Cost ($CAD)
Sodium hydroxide flakes (99%)	T	2,850
Sodium thiosulfate crystals (99%)	T	2,950
Chlorine gas	T	1,750
Process water - fixed cost	quarter year	85
Process water	m3	1.2
Electrical power - fixed cost[3]	Month	12,200
Electrical power - variable cost	MWh	57
Natural gas – fixed cost[4]	Month	185
Natural gas – variable cost	GJ	3.653
Labour – plant manager	person-year	120,000
Labour – process engineer	person-year	96,000
Labour – process operator	person-year	84,000
Labour – lab technician	person-year	78,000

The estimated operating costs for the Demonstration plant are shown in Table 12. The estimated total annual operating cost is $CAD 3.16 million or $CAD 1,753/ t of MgO product. The key operating cost driver is labour, accounting for ~50% of the total. The plant will require 24/7 shift coverage and the number of staff required per shift was based on experience with the analogous plant operating in Spain. On this basis it is believed that there is no opportunity to reduce the total labour numbers. The assumed labour rates were based on current salary data for the region around Rossland, BC.

[3] FORTISBC INC. ELECTRIC TARIFF FOR SERVICE IN THE WEST KOOTENAY AND OKANAGAN AREAS July 1, 2019

[4] FortisBC Energy Inc. Natural Gas Rate Change Effective January 1, 2022

Table 12: Operating Cost Estimate for the Demonstration Plant.

Item	Annual Quantity	Unit	Unit Cost ($CAD)	$CAD/year
Sodium hydroxide	7	t	2850	21,000
Sodium thirosulfate	12	t	2950	34,000
Chlorine	230	t	1750	404,000
Process water	26,280	m3	1.2	32,000
Electrical Power	1,598	MWh	57	238,000
Natural gas	71,136	GJ	3.65	263,000
Labour	16	person-year	87,375	1,398,000
Solid waste disposal	20	t	500	10,000
Product bags	900	ea	15	14,000
Maintenance materials				520,000
General & Administration				222,000
Total Annual Operating Cost ($CAD)				3,156,000
Total Operating Cost ($CAD/ t MgO product)				1,753

21.3 Alternative Lower Capital Cost Demonstration Plant Design

With the objective to lower the total installed capital cost, an alternative Demonstration Plant design was considered. Specifically, the alternative design would use one stirred tank reactor instead of three for the leaching and precipitation stages. This would decrease the capital cost, but would mean that the plant could only be operated in batch mode. In addition, it was considered that the plant would be housed in an existing building instead of constructing a new one. An allowance of $CAD 1.5 million was included in the estimate to account for renovations to an existing building. The revised capital costs are summarized in Figure 11 and indicate that it could be possible to decrease the total installed capital cost by $CAD 6 million to $CAD 21.9 million.

Table 13: Capital Cost Estimate for the Alternative Demonstration Plant Design.

Major Units	**$CAD**
Leaching	2,255,000
Precipitation	1,275,000
Pyrohydrolysis	5,618,000
Tank farm	2,125,000
Balance of plant	418,000
Buildings	1,500,000
Total Direct Capital Cost	13,191,000
Indirect Costs	
EPCM & start-up services	4,386,000
Freight	935,000
Field indirect & first fill	490,000
Total Indirect Capital Cost	5,811,000
Total Direct and Indirect Costs	19,002,000
Contingency (15%)	2,860,000
Total Installed Capital Cost	**21,860,000**

21.4 Capital Cost Estimate – Commercial Plant

The major equipment for the process were sized based on the output from the mass and energy model. This equipment was then costed using quotes from suppliers and in-house data from similar recent projects. The equipment costs were then used to develop factored costs for installation, civils, piping, electrical and controls/instrumentation to develop a total direct cost. Indirect costs and a 20% contingency were added to the direct costs to estimate the total installed capital cost. The capital cost estimate for the Commercial plant is summarized in Table 14. The total installed capital cost is estimated to be $CAD 56.3 million (-25/+40%).

Table 14: Capital Cost Estimate for the Commercial Plant.

Major Units	$CAD
Leaching	6,320,000
Precipitation	5,692,000
Pyrohydrolysis	13,897,000
Tank farm	4,306,000
Balance of plant	2,591,000
Buildings	5,120,000
Total Direct Capital Cost	37,926,000
Indirect Costs	
EPCM & start-up services	5,408,200
Freight	2,305,800
Field indirect & first fill	1,249,000
Total Indirect Capital Cost	8,963,000
Total Direct and Indirect Costs	46,890,000
Contingency (20%)	9,380,000
Total Installed Capital Cost	**56,270,000**

21.5 Operating Cost Estimate – Commercial Plant

The unit cost assumptions used for the operating cost estimate for the Commercial plant are summarized in Table 15. It is noted that the reagent (sodium hydroxide, sodium thiosulfate and chlorine) costs in Table 15 are lower than those in Table 11, reflecting the greater economies of scale for the Commercial plant.

Table 15: Unit Cost Assumptions for Commercial Plant.

Item	Unit	Price ($CAD)
Sodium hydroxide flakes (99%)	t	700
Sodium thiosulfate crystals (99%)	t	800
Chlorine gas	t	500
Process water - fixed cost	quarter year	85
Process water	m3	1.2
Electrical power - fixed cost	month	12,200
Electrical power - variable cost	MWh	57
Natural gas – fixed cost	month	185
Natural gas – variable cost	GJ	3.653
Labour – plant manager	person-year	120,000
Labour – process engineer	person-year	96,000
Labour – process operator	person-year	84,000
Labour – lab technician	person-year	78,000

The estimated operating costs for the commercial plant are summarized in Table 16. The estimated total annual operating cost is $CAD 8.5 million or $CAD 489/ t of MgO product.

Table 16: Operating Cost Estimate for the Commercial Plant.

Item	Annual Quantity	Unit	Unit Cost ($CAD)	$CAD/year
Sodium hydroxide	72	t	700	51,000
Sodium thirosulfate	115	t	800	93,000
Chlorine	2,160	t	500	1,080,000
Process water	262,800	m3	1.2	316,000
Electrical Power	14,904	MWh	57	996,000
Natural gas	684,000	GJ	3.65	2,501,000
Labour	21	person-year	78,002	1,639,000
Solid waste disposal	200	t	500	100,000
Product bags	8,640	ea	15	130,000
Maintenance materials				1,138,000
General & Administration				410,000
Total Annual Operating Cost ($CAD)				8,454,000
Total Operating Cost ($CAD/ t MgO product)				489

21.6 Commercial Caustic Calcined Magnesia Plant

The baseline Commercial plant strategy analyzed in this report is to produce a 98% spray roasted MgO powder product. With additional processing steps it would be possible to take this product and produce a caustic calcined magnesia (CCM) with a purity >99% with the potential for a higher selling price. As illustrated in Figure 18, it would require the addition of "MgO Washing" followed by "Calcination" in a multi-hearth furnace.

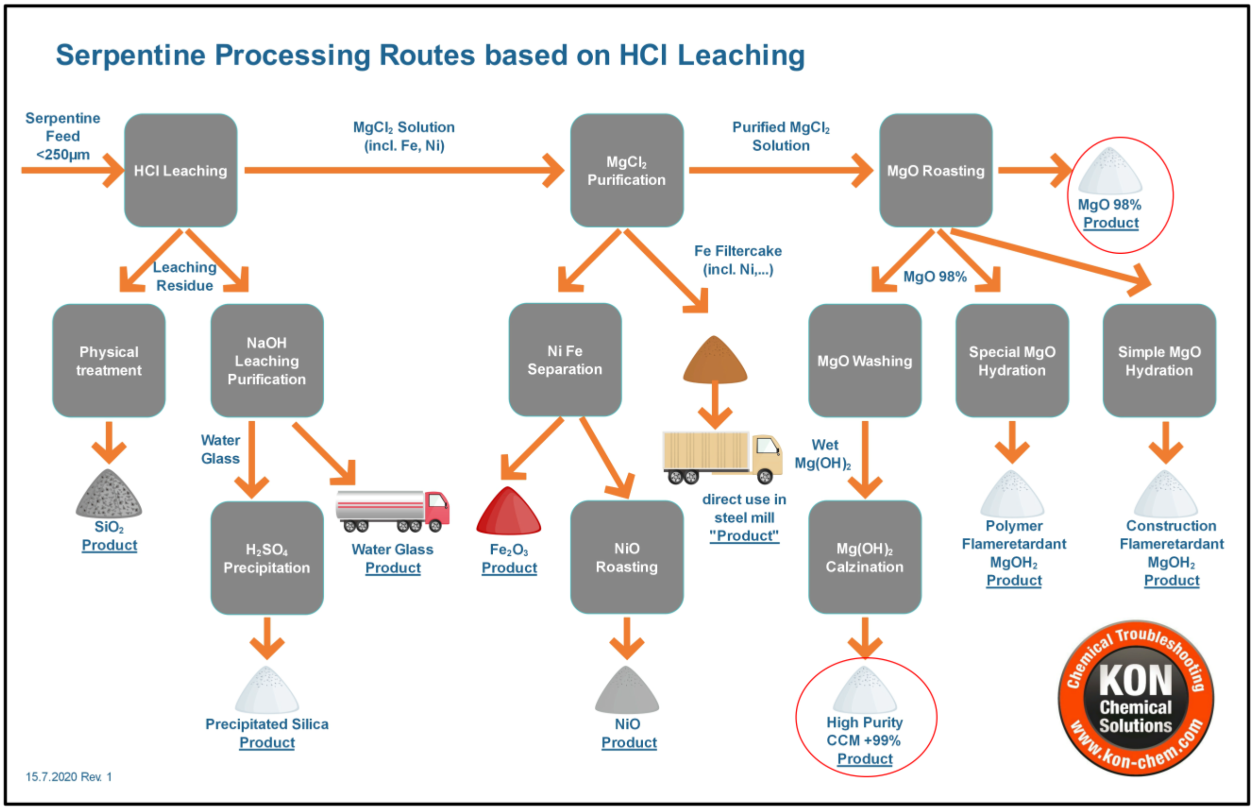


Figure 18: Potential Processing Routes.

21.6.1High Level Economics of CCM Plant

High-level estimates of the capital and operating costs for a commercial plant producing >99% CCM, based on the design and unit cost data in Sections 17.3.1-21.5, are shown in Table 17 and Table 18, respectively.

In summary, it is estimated that the addition of MgO washing and Calcination to produce a CCM (>99% MgO) product would add $CAD 35 million to the plant TIC for a total of $CAD 91.2 million. It was estimated that this plant would produce approximately 17,300 t MgO annually. The additional labour, natural gas, electrical power, and maintenance required to produce CCM would increase the total annual operating cost to ~ $CAD 12.5 million annually or $CAD 721/t product, compared to $CAD 489/t for the spray roasted (98% MgO) product.

Table 17: Capital Cost Estimate for a CCM Commercial Plant.

Major Units	**$CAD**
Leaching	6,607,000
Precipitation	5,951,000
Pyrohydrolysis	14,529,000
Tank farm	4,502,000
MgO washing & calcination	17,774,000
Balance of plant	4,233,000
Buildings	8,259,000
Total Direct Capital Cost	61,855,000
Indirect Costs	
EPCM & start-up services	7,994,000
Freight	4,060,000
Field indirect	1,624,000
First fill	460,000
Total Indirect Capital Cost	14,138,000
Total Direct and Indirect Costs	75,990,000
Contingency (20%)	15,200,000
Total Installed Capital Cost	**91,190,000**

Item	**Annual Quantity**	**Unit**	**Unit Cost ($CAD)**	**$CAD/year**
Sodium hydroxide	72	t	700	51,000
Sodium thirosulfate	115	t	800	93,000
Chlorine	2,160	t	500	1,080,000
Process water	646,560	m3	1.2	777,000
Electrical Power	40,176	MWh	57	2,437,000
Natural gas	926,957	GJ	3.65	3,389,000
Labour	25	person-year	78,005	1,951,000
Solid waste disposal	200	t	500	100,000
Product bags	8,640	ea	15	130,000
Maintenance materials				1,856,000
General & Administration				600,000
			Total Annual Operating Cost ($CAD)	12,464,000
			Total Operating Cost ($CAD/ t MgO product)	721

Table 18: Operating Cost Estimate for a CCM Commercial Plant.

22 ECONOMIC ANALYSIS

22.1 Introduction

WHY's objective is to operate a plant treating 250,000 t/y ore from the Record Ridge deposit. For this study, it was assumed that the commercial plant would be based on installation of 5 processing modules of 50,000 t/y (each called a unit).

An engineering economic model was prepared for the Project to estimate annual cash flows and assess sensitivities to certain economic parameters. The economic results of this report are based upon the PFS performed by KPM in collaboration with Tenova and KON Chemical Solutions e.U in 2022 and the PEA of SRK in 2013.

22.2 Financial Model Parameters

A base case MgO price of US$1,500/t is based on consensus sponsor estimates. The forecasts are meant to reflect the average MgO price expectation over the life of the project. No price inflation or escalation factors were considered. Commodity prices can be volatile, and there is the potential for deviation from the forecast. The economic analysis was performed using the following assumptions:

- Commercial plant of 250,000 t/a ore capacity was considered on modular basis of five x 50,000 t/a units, with no consideration for CAPEX/OPEX savings for larger plants;
- Construction starting January 1, 2023;
- All construction costs capitalized in Year -1;
- Commercial production starting on January 1, 2024, with no production ramp up, first revenue and expensed costs in Year +1;
- Project life of 20 years;
- An exchange rate of US$0.73 per C$1.00 was assumed;
- Cost estimates in constant Q1 2022 American dollars with no inflation or escalation;
- 100% ownership;
- Capital costs funded with 50% equity (no financing costs assumed) and 50% debt (8% interest);
- MgO is assumed to be sold in the same year it is produced;
- No contractual arrangements currently exist;

Taxes

The project has been evaluated on an after-tax basis to provide an approximate value of the potential economics. The project was assumed to be subject to the Canadian corporate income tax system which consists of 12% federal income tax.

Working Capital

Working capital and initial fills are not included in the valuation. The effective sum of working capital and initial fills over the life of the project is zero.

Closure Costs & Salvage Value

Neither project closure cost nor a salvage value have been considered

22.3 Economic Analysis

The economic analysis was performed for a commercial plant of 250,000 t/y throughput assuming a 5% discount rate. The after- tax NPV discounted at 5% is US$872 M; the internal rate of return IRR is 72.0%; and payback period is 1.43 years. A summary of project economics is shown below.

Project Assumptions	
Project duration, yr	20
Required Mining tonnage per unit, t ore	50,000
Operating costs per unit, $/t MgO	375 $
Production Units	5
MgO sales price, $/t	1 500 $
Loan Interest	8,00%
Loan Duration, months	120
Debt portion	50,00%
Straight line method Depreciation, yr	10
Capital Expenditure per Unit	41 076 370 $
Plant construction loan	102 690 925 $
Sustaining Capital	0 $
Net Working Capital	0 $
Salvage Value of assets	0 $
Book Value of assets	0 $
Income Tax Rate	12,00%
Discount Rate (WACC)	5,00%
Opportunity cost	0 $
Closure cost	0 $

Business Results	Project Value	Conditions	Decision
NPV of Cash Flow	$ 871 774 903	>0	Yes
IRR	72,0%	> 5%	Yes
Simple Payback	1,43	< 5	Yes
Discounted Paybac	1,50	< 5	Yes
Profitability Index	15,50	> 4	Yes

22.4 Methodology Used

An economic model was developed to estimate annual post-tax cash flows and sensitivities of the project based on a 5% discount rate. A sensitivity analysis was performed to assess the impact of variations in magnesium prices, total operating cost, capital expenditures, and discount rate.

The capital and operating cost estimates developed specifically for this project were given by KPM and Tenova. The economic analysis has been run on a constant dollar basis with no inflation. Due to an uncertainty of timing cash outflows during the pre-production period, a simplified 12-month construction period was used which includes all pre-production capital expenditures.

22.5 Cautionary Statement

The results of the economic analyses represent forward-looking information as defined under Canadian securities law. The results depend on inputs that are subject to several known and unknown risks, uncertainties, and other factors that may cause actual results to differ materially from those presented herein. Information that is forward-looking includes the following:

- Mined and delivered magnesium ore availability for all the duration of the project;
- Proposed commodity prices and exchange rates;
- Proposed plant production plan;
- Projected process recovery rates;
- Proposed operating costs;
- Proposed capital expenditures;
- Sustaining capital expenditures;
- Closure costs and closure requirements;
- Change in net working capital;
- Salvage value of assets;
- Resale value and book value;
- Assumptions about environmental, permitting, and social risks;

22.6 Principal Assumptions

The cash flow estimate includes only revenue, costs, interest rate for the loan, income tax, and other factors applicable to the Project. Corporate obligations, financing costs, and other expenses at the corporate level are excluded.

The model was prepared from mining schedules estimated on an annual basis. The cash flow model was based on the following:

- All costs are reported in American dollars (US$) and referenced as '$', unless otherwise stated.
- Fifty percent (50%) equity basis.
- No cost escalation beyond 2022.
- No provision for effects of inflation.
- Constant 2022-dollar analysis.
- The economic analysis consists of the technical assumptions outlined in the report of KPM section 3.1 and with the economic assumptions and estimated Capital and Operating costs described in this report (sections 3.4 and 3.5 convert in USD, 1 CAD = 0.73 USD).

Table 22.1. Capital Cost estimate for the Commercial Plant unit of 50,000 t/y ore.

Major Units	**$US**	
Leaching	$	4 613 600
Precipitation	$	4 155 160
Pyrohydrolysis	$	10 144 810
Tank farm	$	3 143 380
Balance of plant	$	1 891 430
Buildings	$	3 737 600
Total Direct Capital Cost	**$**	**27 685 980**
Indirect Costs		
EPCM & start-up services	$	3 947 986
Freight	$	1 683 234
Field Indirect Capital Cost	$	911 770
Total Inderect Capital Costs	**$**	**6 542 990**
Total Direct and Indirect Costs	**$**	**34 228 970**
Contingency (20%)	**$**	**6 847 400**
Total Installed Capital Cost	**$**	**41 076 370**

Table 22.2. Operating Cost Estimate for the Commercial Plant unit of 50,000 t/y ore.

Item	Annual Quantity	Unit		$US/year
Sodium Hydroxide	72	t		$ 37 230
Sodium thirosulfate	115	t		$ 67 890
Chlorine	2 160	t		$ 788 400
Process water	262 800	m3		$ 230 680
Electrical Power	14 904	MWh		$ 727 080
Natural Gas	684 000	GJ		$ 1 825 730
Labour	21	person-year		$ 1 196 470
Solid waste disposal	200	t		$ 73 000
Product bags	8 640	ea		$ 94 900
Mining	17 288	t MgO		$ 315 506
Maintenance materials				$ 830 740
General & Administration				$ 299 300
		Total Annuel Operating Cost ($US)		**$ 6 486 926**
		Total Operating and mining Costs ($US/ t MgO product)		**$ 375**

22.7 Cashflow Model, Base Case, MgO 5-unit commercial plant

The results are derived from the life of mine of 20 years schedule, and Capex and Opex are represented in Tables 22.1 and 22.2. The Table 22.3 shows the Cashflow model results.

Table 22.3. Cashflow Model, Base Case. Post-Tax NPV & IRR.

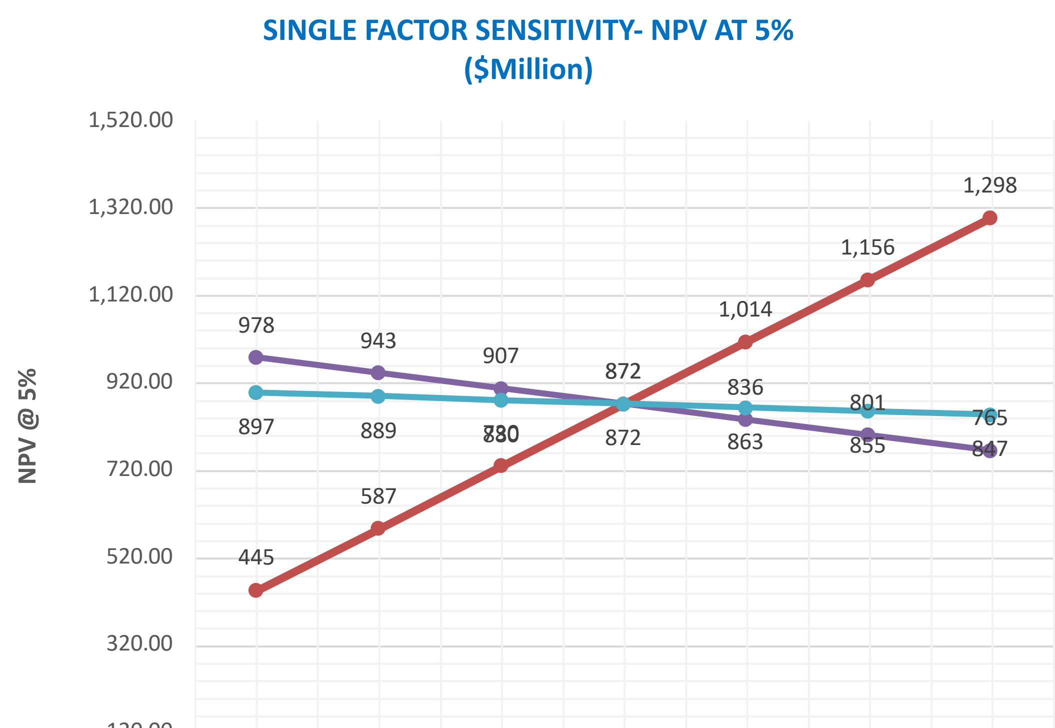
Financial analysis (WHY project ($US)) - MgO 5 units																
			-1	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11-20**	**Total**
			2 023	**2 024**	**2 025**	**2 026**	**2 027**	**2 028**	**2 029**	**2 030**	**2 031**	**2 032**	**2 033**	**2 043**	
Investment cost (CAPEX,$US)			205 381 850												
Debt portion		50%	102 690 925												
Plant	Capital at the end			102 690 925	95 702 556	88 134 156	79 937 582	71 060 696	61 447 034	51 035 442	39 759 694	27 548 063	14 322 874	0	
	Capital payment	10		6 988 369	7 568 400	8 196 574	8 876 885	9 613 662	10 411 592	11 275 749	12 211 630	13 225 190	14 322 874	0	
Production	**Total MgO productior**	**t**		**86 440**	**86 440**	**86 440**	**86 440**	**86 440**	**86 440**	**86 440**	**86 440**	**86 440**	**86 440**	**86 440**	**1 728 800**
Revenues		Price	1 500												
Total revenues	**Total revenues**		**2 593 200 000**	**129 660 000**	**129 660 000**	**129 660 000**	**129 660 000**	**129 660 000**	**129 660 000**	**129 660 000**	**129 660 000**	**129 660 000**	**129 660 000**	**129 660 000**	**2 593 200 000**
	Total Opex		**648 692 600**	32 434 630	32 434 630	32 434 630	32 434 630	32 434 630	32 434 630	32 434 630	32 434 630	32 434 630	32 434 630	32 434 630	648 692 600
	Total amortization	10	**205 381 850**	20 538 185	20 538 185	20 538 185	20 538 185	20 538 185	20 538 185	20 538 185	20 538 185	20 538 185	20 538 185		205 381 850
	Total interest	8%	**46 819 990**	7 962 722	7 382 691	6 754 518	6 074 206	5 337 429	4 539 500	3 675 343	2 739 461	1 725 902	628 218	0	46 819 990
Expenses	**Total expenses**		**900 894 440**	**60 935 537**	**60 355 506**	**59 727 333**	**59 047 021**	**58 310 244**	**57 512 315**	**56 648 158**	**55 712 276**	**54 698 717**	**53 601 033**	**32 434 630**	**900 894 440**
Profit before income tax	**EBIT**	**USD**	**1 692 305 560**	**68 724 463**	**69 304 494**	**69 932 667**	**70 612 979**	**71 349 756**	**72 147 685**	**73 011 842**	**73 947 724**	**74 961 283**	**76 058 967**	**97 225 370**	**1 692 305 560**
Income tax		12%	**203 076 667**	8 246 936	8 316 539	8 391 920	8 473 557	8 561 971	8 657 722	8 761 421	8 873 727	8 995 354	9 127 076	11 667 044	203 076 667
Net profit				**60 477 527**	**60 987 955**	**61 540 747**	**62 139 421**	**62 787 785**	**63 489 963**	**64 250 421**	**65 073 997**	**65 965 929**	**66 931 891**	**85 558 326**	**1 489 228 893**
Amortization				20 538 185	20 538 185	20 538 185	20 538 185	20 538 185	20 538 185	20 538 185	20 538 185	20 538 185	20 538 185	0	205 381 850
Total capital repayments				6 988 369	7 568 400	8 196 574	8 876 885	9 613 662	10 411 592	11 275 749	12 211 630	13 225 190	14 322 874	0	102 690 925
After Tax Cashflow			**-102 690 925**	**74 027 343**	**73 957 739**	**73 882 358**	**73 800 721**	**73 712 308**	**73 616 556**	**73 512 857**	**73 400 552**	**73 278 925**	**73 147 202**	**85 558 326**	**1 489 228 893**
Discounted Cash Flow			**-102 690 925**	**70 502 231**	**67 081 850**	**63 822 359**	**60 716 036**	**57 755 522**	**54 933 808**	**52 244 215**	**49 680 383**	**47 236 248**	**44 906 037**	**32 246 033**	**871 774 903**
Cumulated cashflow			-102 690 925	-28 663 582	45 294 157	119 176 516	192 977 237	266 689 545	340 306 101	413 818 958	487 219 510	560 498 435	633 645 637	1 489 228 893	
Discounted cumulated cashflow			-102 690 925	-32 188 694	34 893 156	98 715 515	159 431 551	217 187 073	272 120 881	324 365 096	374 045 479	421 281 727	466 187 764	871 774 903	
Simple Pay back Period			**1,43**	1,00	0,43										
Discounted Pay Back Period			**1,50**	1,00	0,50										
IRR			**72,03%**												
NPV			**871 774 903 $**												
Profitabilty Index			**15,50**												

Table 22.4. Cashflow Model, Base Case. Pre-Tax NPV & IRR

Financial analysis (WHY project ($US)) - MgO 5 units															
			-1	1	2	3	4	5	6	7	8	9	10	11-20	Total
			2 023	2 024	2 025	2 026	2 027	2 028	2 029	2 030	2 031	2 032	2 033	2 043	
Investment cost (CAPEX,$US)			205 381 850												
Debt portion		50%	102 690 925												
Plant	Capital at the end			102 690 925	95 702 556	88 134 156	79 937 582	71 060 696	61 447 034	51 035 442	39 759 694	27 548 063	14 322 874	0	
	Capital payment	10		6 988 369	7 568 400	8 196 574	8 876 885	9 613 662	10 411 592	11 275 749	12 211 630	13 225 190	14 322 874	0	
Production	Total MgO productior	t		86 440	86 440	86 440	86 440	86 440	86 440	86 440	86 440	86 440	86 440	86 440	1 728 800
Revenues		Price	1 500												
Total revenues	Total revenues		2 593 200 000	129 660 000	129 660 000	129 660 000	129 660 000	129 660 000	129 660 000	129 660 000	129 660 000	129 660 000	129 660 000	129 660 000	2 593 200 000
	Total Opex		**648 692 600**	32 434 630	32 434 630	32 434 630	32 434 630	32 434 630	32 434 630	32 434 630	32 434 630	32 434 630	32 434 630	32 434 630	648 692 600
	Total amortization	10	**205 381 850**	20 538 185	20 538 185	20 538 185	20 538 185	20 538 185	20 538 185	20 538 185	20 538 185	20 538 185	20 538 185		205 381 850
	Total interest	8%	**46 819 990**	7 962 722	7 382 691	6 754 518	6 074 206	5 337 429	4 539 500	3 675 343	2 739 461	1 725 902	628 218	0	46 819 990
Expenses	Total expenses		900 894 440	60 935 537	60 355 506	59 727 333	59 047 021	58 310 244	57 512 315	56 648 158	55 712 276	54 698 717	53 601 033	32 434 630	900 894 440
Profit before income tax	EBIT	USD	1 692 305 560	68 724 463	69 304 494	69 932 667	70 612 979	71 349 756	72 147 685	73 011 842	73 947 724	74 961 283	76 058 967	97 225 370	1 692 305 560
Income tax		0%	**0**	0	0	0	0	0	0	0	0	0	0	0	0
Net profit				68 724 463	69 304 494	69 932 667	70 612 979	71 349 756	72 147 685	73 011 842	73 947 724	74 961 283	76 058 967	97 225 370	1 692 305 560
Amortization				20 538 185	20 538 185	20 538 185	20 538 185	20 538 185	20 538 185	20 538 185	20 538 185	20 538 185	20 538 185	0	205 381 850
Total capital repayments				6 988 369	7 568 400	8 196 574	8 876 885	9 613 662	10 411 592	11 275 749	12 211 630	13 225 190	14 322 874	0	102 690 925
After Tax Cashflow			-102 690 925	82 274 279	82 274 279	82 274 279	82 274 279	82 274 279	82 274 279	82 274 279	82 274 279	82 274 279	82 274 279	97 225 370	1 692 305 560
Discounted Cash Flow			-102 690 925	78 356 456	74 625 196	71 071 615	67 687 253	64 464 050	61 394 333	58 470 794	55 686 470	53 034 734	50 509 270	36 643 219	993 503 721
Cumulated cashflow			-102 690 925	-20 416 646	61 857 632	144 131 911	226 406 189	308 680 468	390 954 746	473 229 025	555 503 303	637 777 582	720 051 860	1 692 305 560	
Discounted cumulated cashflow			-102 690 925	-24 334 469	50 290 727	121 362 342	189 049 594	253 513 644	314 907 978	373 378 772	429 065 242	482 099 975	532 609 245	993 503 721	
Simple Pay back Period			1,27	1,00	0,27										
Discounted Pay Back Period			1,34	1,00	0,34										
IRR			80,16%												
NPV			993 503 721 $												
Profitabilty Index			17,48												

22.8 Sensitivity Analysis

The single factor sensitivity at a 5% discount shows that commodity prices are the largest single uncertainty with respect to project value.



	-30%	-20%	-10%	Base	10%	20%	30%
MgO Price	445	587	730	872	1014	1156	1298
Opex	978	943	907	872	836	801	765
Capex	897	889	880	872	863	855	847

A Two-factor sensitivity – Price and Discount rate shows a positive valuation is maintained across a wide range of sensitivities on key assumptions such as MgO prices and discount rate.

Two-factor sensitivity (NPV in $M) – Price and Discount Rate							
871 774 903 $	**1 050 $**	**1 200 $**	**1 350 $**	**1 500 $**	**1 650 $**	**1 800 $**	**1 950 $**
3,50%	527 859 780 $	690 024 438 $	852 189 097 $	**1 014 353 756 $**	1 176 518 414 $	1 338 683 073 $	1 500 847 732 $
4,00%	498 407 986 $	653 474 697 $	808 541 408 $	**963 608 119 $**	1 118 674 830 $	1 273 741 540 $	1 428 808 251 $
4,50%	470 901 899 $	619 323 495 $	767 745 090 $	**916 166 686 $**	1 064 588 282 $	1 213 009 877 $	1 361 431 473 $
5,00%	**445 190 452 $**	**587 385 269 $**	**729 580 086 $**	**871 774 903 $**	**1 013 969 720 $**	**1 156 164 537 $**	**1 298 359 354 $**
5,50%	421 135 594 $	557 490 414 $	693 845 234 $	**830 200 054 $**	966 554 874 $	1 102 909 695 $	1 239 264 515 $
6,00%	398 611 079 $	529 483 798 $	660 356 517 $	**791 229 235 $**	922 101 954 $	1 052 974 673 $	1 183 847 391 $
6,50%	377 501 374 $	503 223 423 $	628 945 473 $	**754 667 522 $**	880 389 571 $	1 006 111 620 $	1 131 833 669 $

22.9 Cashflow Model, MgO 1 unit of 50,000 t/y ore.

The following cashflow model shows the results for MgO product with one unit of production.

Table 22.5. Cashflow Model, MgO product, 1 unit of production.

Financial analysis (WHY project ($US)) - MgO 1 unit															
			-1	1	2	3	4	5	6	7	8	9	10	11-20	Total
			2 023	2 024	2 025	2 026	2 027	2 028	2 029	2 030	2 031	2 032	2 033	2 043	
Investment cost (CAPEX,$US)			41 076 370												
Debt portion		50%	20 538 185												
Plant	Capital at the end			20 538 185	19 140 511	17 626 831	15 987 516	14 212 139	12 289 407	10 207 088	7 951 939	5 509 613	2 864 575	0	
	Capital payment	10		1 397 674	1 513 680	1 639 315	1 775 377	1 922 732	2 082 318	2 255 150	2 442 326	2 645 038	2 864 575	0	
Production	Total MgO productior	t		17 288	17 288	17 288	17 288	17 288	17 288	17 288	17 288	17 288	17 288	17 288	345 760
Revenues		Price	1 500												
Total revenues	Total revenues		518 640 000	25 932 000	25 932 000	25 932 000	25 932 000	25 932 000	25 932 000	25 932 000	25 932 000	25 932 000	25 932 000	25 932 000	518 640 000
	Total Opex		129 738 520	6 486 926	6 486 926	6 486 926	6 486 926	6 486 926	6 486 926	6 486 926	6 486 926	6 486 926	6 486 926	6 486 926	129 738 520
	Total amortization	10	41 076 370	4 107 637	4 107 637	4 107 637	4 107 637	4 107 637	4 107 637	4 107 637	4 107 637	4 107 637	4 107 637	0	41 076 370
	Total interest	8%	9 363 998	1 592 544	1 476 538	1 350 904	1 214 841	1 067 486	907 900	735 069	547 892	345 180	125 644	0	9 363 998
Expenses	Total expenses		115 309 628	12 187 107	12 071 101	11 945 467	11 809 404	11 662 049	11 502 463	11 329 632	11 142 455	10 939 743	10 720 207	6 486 926	180 178 888
Profit before income tax	EBIT	USD	338 461 112	13 744 893	13 860 899	13 986 533	14 122 596	14 269 951	14 429 537	14 602 368	14 789 545	14 992 257	15 211 793	19 445 074	338 461 112
Income tax		12%	40 615 333	1 649 387	1 663 308	1 678 384	1 694 711	1 712 394	1 731 544	1 752 284	1 774 745	1 799 071	1 825 415	2 333 409	40 615 333
Net profit				12 095 505	12 197 591	12 308 149	12 427 884	12 557 557	12 697 993	12 850 084	13 014 799	13 193 186	13 386 378	17 111 665	297 845 779
Amortization				4 107 637	4 107 637	4 107 637	4 107 637	4 107 637	4 107 637	4 107 637	4 107 637	4 107 637	4 107 637	0	41 076 370
Total capital repayments				1 397 674	1 513 680	1 639 315	1 775 377	1 922 732	2 082 318	2 255 150	2 442 326	2 645 038	2 864 575	0	20 538 185
After Tax Cashflow			-20 538 185	14 805 469	14 791 548	14 776 472	14 760 144	14 742 462	14 723 311	14 702 571	14 680 110	14 655 785	14 629 440	17 111 665	297 845 779
Discounted Cash Flow			-20 538 185	14 100 446	13 416 370	12 764 472	12 143 207	11 551 104	10 986 762	10 448 843	9 936 077	9 447 250	8 981 207	6 449 207	174 354 981
Cumulated cashflow			-20 538 185	-5 732 716	9 058 831	23 835 303	38 595 447	53 337 909	68 061 220	82 763 792	97 443 902	112 099 687	126 729 127	297 845 779	
Discounted cumulated cashflow			-20 538 185	-6 437 739	6 978 631	19 743 103	31 886 310	43 437 415	54 424 176	64 873 019	74 809 096	84 256 345	93 237 553	174 354 981	
Simple Pay back Period			1,43	1,00	0,43										
Discounted Pay Back Period			1,50	1,00	0,50										
IRR			72,0%												
NPV			174 354 981 $												
Profitabilty Index			15,50												

22.10 Cashflow Model, CCM 1 unit of 50,000 t/y ore.

The following cashflow model shows the results for CCM product with one unit of production.

Table 22.6. Cashflow Model, CCM product, 1 unit of production.

Financial analysis (WHY project ($US)) - CCM 1 unit															
			-1	1	2	3	4	5	6	7	8	9	10	11-20	Total
			2 023	2 024	2 025	2 026	2 027	2 028	2 029	2 030	2 031	2 032	2 033	2 043	
Investment cost (CAPEX,$US)			66 570 890												
Debt portion		50%	33 285 445												
Plant	Capital at the end			33 285 445	31 020 289	28 567 126	25 910 352	23 033 066	19 916 968	16 542 235	12 887 401	8 929 217	4 642 506	0	
	Capital payment	10		2 265 156	2 453 163	2 656 774	2 877 285	3 116 098	3 374 733	3 654 834	3 958 184	4 286 711	4 642 506	0	
Production	Total MgO production	t		17 288	17 288	17 288	17 288	17 288	17 288	17 288	17 288	17 288	17 288	17 288	345 760
Revenues		Price	2 200												
Total revenues	Total revenues		760 672 000	38 033 600	38 033 600	38 033 600	38 033 600	38 033 600	38 033 600	38 033 600	38 033 600	38 033 600	38 033 600	38 033 600	760 672 000
	Total Opex		**188 284 520**	9 414 226	9 414 226	9 414 226	9 414 226	9 414 226	9 414 226	9 414 226	9 414 226	9 414 226	9 414 226	9 414 226	188 284 520
	Total amortization	10	**66 570 890**	6 657 089	6 657 089	6 657 089	6 657 089	6 657 089	6 657 089	6 657 089	6 657 089	6 657 089	6 657 089	0	66 570 890
	Total interest	8%	**15 175 871**	2 580 975	2 392 969	2 189 357	1 968 846	1 730 033	1 471 398	1 191 297	887 948	559 421	203 626	0	15 175 871
Expenses	Total expenses		270 031 281	18 652 290	18 464 284	18 260 672	18 040 161	17 801 348	17 542 713	17 262 612	16 959 263	16 630 736	16 274 941	9 414 226	270 031 281
Profit before income tax	EBIT	USD	490 640 719	19 381 310	19 569 316	19 772 928	19 993 439	20 232 252	20 490 887	20 770 988	21 074 337	21 402 864	21 758 659	28 619 374	490 640 719
Income tax		12%	**58 876 886**	2 325 757	2 348 318	2 372 751	2 399 213	2 427 870	2 458 906	2 492 519	2 528 920	2 568 344	2 611 039	3 434 325	58 876 886
Net profit				17 055 552	17 220 998	17 400 176	17 594 226	17 804 382	18 031 980	18 278 469	18 545 417	18 834 521	19 147 620	25 185 049	431 763 833
Amortization				6 657 089	6 657 089	6 657 089	6 657 089	6 657 089	6 657 089	6 657 089	6 657 089	6 657 089	6 657 089	0	66 570 890
Total capital repayments				2 265 156	2 453 163	2 656 774	2 877 285	3 116 098	3 374 733	3 654 834	3 958 184	4 286 711	4 642 506	0	33 285 445
After Tax Cashflow			-33 285 445	21 447 485	21 424 924	21 400 491	21 374 030	21 345 372	21 314 336	21 280 724	21 244 322	21 204 899	21 162 203	25 185 049	431 763 833
Discounted Cash Flow			-33 285 445	20 426 176	19 433 038	18 486 549	17 584 467	16 724 658	15 905 086	15 123 813	14 378 993	13 668 867	12 991 757	9 491 980	250 827 066
Cumulated cashflow			-33 285 445	-11 837 960	9 586 965	30 987 456	52 361 486	73 706 858	95 021 194	116 301 918	137 546 240	158 751 139	179 913 342	431 763 833	
Discounted cumulated cashflow			-33 285 445	-12 859 269	6 573 770	25 060 318	42 644 786	59 369 443	75 274 529	90 398 342	104 777 336	118 446 202	131 437 959	250 827 066	
Simple Pay back Period			**1,61**	1,00	0,61										
Discounted Pay Back Period			**1,70**	1,00	0,70										
IRR			**64,4%**												
NPV			**250 827 066 $**												
Profitabilty Index			**13,97**												