D6.4 Blueprint and techno-economic analysis of the NMC battery recycling processes

Project	Green Transport Delta - Electrification
Project number	MOB21011
Author(s)	M. Jovicic with contribution of E. Folkertsma, S. van Berkum, S. van Wyk, M. Roelands
Version	06
Date	22 October 2024

Revision index

Version	Description	Date
00	Draft version of the report	04.06.2024
01	Changed structure of the report from more scientific to more decision maker-oriented report	20.06.2024
02	Aligned mixed precipitation and solvent extraction assumptions	12.07.2024
03	Added main results in figures and tables	03.08.2024
04	First review and correction	06.09.2024
05	Second review and correction	10.10.2024
06	Final review and correction	18.10.2024

Contact details

Name	Role	Email
Milkica Jovicic	Technical consultant	milkica.jovicic@tno.nl
Emma Folkertsma	Project manager and work package 6 lead	emma.folkertsma@tno.nl

The Green Transport Delta – Electrification project has received financial support from the Ministry of Economic Affairs and Climate, under the grant ‘R&D Mobility Sectors’ carried out by the Netherlands Enterprise Agency.

Content

Executive summary

The Green Transport Delta Electrification project is boosting the knowledge and export position of the Dutch industry in the domain of electrification of the mobility system. The “Battery recycling” part focusses on the development of **knowledge and practical experience** with recovery of valuable materials from end-of-life lithium ion batteries by investigating:

- current and future battery compositions,
- suitability, performance, and in- and output materials for recycling of various materials via metallurgical separation methods,
- separation techniques of black mass components.

The main objective of the work presented here is to further build on the earlier published report *Literature assessment of Li-ion battery recycling steps*. This is done by the development of a perspective on a scalable and flexible battery recycling process by **setting up a blueprint for battery recycling** and evaluate it through the preliminary techno-economic assessment. The development of the methodology used in this work was a major goal of this project, with the benefit that it is also suitable to apply for alternative metal recycling processes or new iterations of techno-economic assessment with better specified information.

Building of the blueprint is done in two phases – **construction phase** where the battery limits for the recycling processes are set and input data is collected, and **utilization phase** which defined process schemes for two battery recycling routes (mixed precipitation and solvent extraction), their main energy and material flows and services. The blueprint is then used as a basis for the preliminary techno-economic assessment. The results presented here are all based on the limited information publicly available from patents on the corresponding processes. For this work, several assumptions on process details have been made, as mentioned in this report.

The developed blueprint provides a comprehensive representation of **two industrial process routes** (mixed precipitation and solvent extraction) while it describes their major challenges and opportunities. Both processes differ on process complexity, use of auxiliary chemicals, final products and their value, and eventually on the field of application. Based on the preliminary techno-economic assessment conducted for these two routes, it can be concluded that mixed precipitation route is a more environmentally friendly process due to less solvent use, with lower process complexity and lower investment costs than the solvent extraction route. Both process routes result in the formation of considerable amounts of waste salt brines for which discharge is becoming more restricted. The mixed precipitation route’s main drawback is that a final product is obtained in the form of a mixed precipitate nickel, cobalt, and manganese hydroxide cake, while a benefit of the solvent recovery route are potentially higher revenues because these critical materials are recovered as separate final products.

During the execution of this task, gaps were identified in terms of missing data on non-clear process operations. A two-way approach is recommended to match the envisioned objective and the practice of working out the task: top-down and bottom-up analysis of these processes. Top-down analysis would serve to identify and solve potential issues in the battery recycling value chain, while the bottom-up analysis would follow the similar line like this report, but it would be based on experimental data in order to minimize number of assumptions on battery composition and related process flexibility, process conditions, solvent use, solvent choice, etc. In this way, both types of analysis can benefit from each other and cover the most important questions at the early stage of setting a national battery recycling value chain.

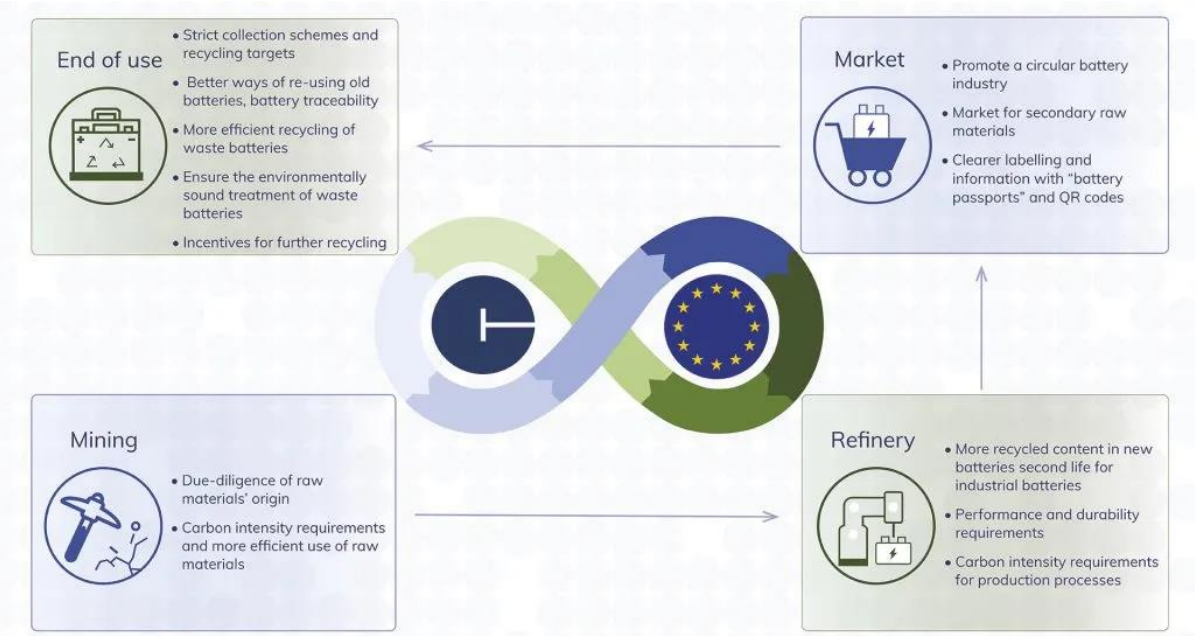
1. Introduction

1.1 Background

The European Union (EU) has set ambitious targets to reduce carbon emissions and increase the use of sustainable energy. This had a domino effect as the demand for renewable energy systems and electric vehicles increased, and it is possible to foresee the growth in the battery manufacturing and recycling industry with a key focus on lithium-ion batteries (LIB). While identifying batteries as a key technology to drive the green transition (European Commission, 2023), the EU set as a strategic imperative becoming a leader in the battery industry by supporting the creation of sustainable batteries and competitive value chains in Europe. The EU has granted projects worth 430 MEUR for research and innovation to support the value chain to strengthen the industrial leadership potential of the EU battery industry, and more funding is projected for the future. The European battery industry is expected to grow at a fast pace from 10 GWh in 2020 to more than 400 GWh in 2030 (Blackridge Research & Consulting, 2024).

The new EU Battery Regulation (European Parliament, 2023) entered into force in August 2023 and is in compliance with the European Green Deal (The European Green Deal, 2024), and will ensure batteries are collected, reused and recycled to the highest degree to lower the carbon footprint, improve sustainability and competitiveness of battery economy. The Regulation sets new requirements for battery design, manufacture, labelling, recycling and reporting as it aims to promote safety, energy and material efficiency, long life and environmental friendliness of batteries by reducing health hazards.

Battery recycling is recognized as one of the main challenges in EU due to the diversity in battery composition and scarcity of its raw materials. In the Netherlands, there are currently no companies that retrieve the critical raw materials from high capacity batteries from the mobility sector, but there are some companies who do collection, sorting and pretreatment. The opportunity for the Netherlands to become a significant player in battery recycling value chain, with a specific focus on nickel-cobalt-manganese (NMC) black mass recycling was a motive to set the path starting from a blueprint followed by the first-order techno-economic assessment. The value chain of batteries is shown in Figure 1.



Source: Council of the European Union, European Commission
© Celltech, 2023

Figure 1: *Value chain of batteries* (European Parliament, 2023)

1.2 Aim of the report

The aim of this document is to give a comprehensive representation of two hydrometallurgical battery recycling routes, their main energy and material flows, and services, followed by a preliminary techno-economic assessment and sensitivity analysis. The results presented here are all based on the limited information that is publicly available from patents on the corresponding processes, consequently it was inevitable to make several assumptions for the work to be done. The development of the methodology used in this work was an important goal of this project, because this method is also suitable to apply to alternative metal recycling processes or to new iterations of the studied processes with better specified information.

Battery recycling was previously explained in the published report *Literature assessment of Li-ion battery recycling steps* (M. B. Rossi, 2024). Black mass is an intermediate product in the battery recycling value chain: batteries are discharged and dismantled prior to a shredding process step. This process step is performed to liberate the target materials: cobalt, nickel, manganese, and lithium. Subsequently, the target materials are separated by state-of-the-art separation processes, yielding black mass, the material stream containing the target materials. Several processing routes for black mass recycling have been developed, however, according to the patents and different scientific papers, two hydrometallurgical routes stand out: black mass recycling using the mixed precipitation route, which ends up in obtaining the nickel-manganese-cobalt (NMC) cake (in the form of hydroxide) and lithium precipitate (in our study case, as a carbonate Li_2CO_3) and black mass recycling using the solvent extraction route to retrieve the individual critical materials (in this study case nickel, manganese and cobalt as sulphates, and lithium as a carbonate).

This report focuses on the techno-economic assessment for the NMC battery recycling process or, more precisely, black mass recycling and makes a high-over comparison of both black mass recycling routes. The first step in this work was to define a blueprint through two phases – data collection (construction phase) and data processing (utilization phase) where the system boundaries are defined. The second step was to perform a techno-economic assessment for the defined system boundaries followed by a sensitivity analysis.

1.3 Intended audience

The intended audience of this report are current project partners, decision makers, and all interested stakeholders from the battery manufacturing and recycling value chain that need an insight into main challenges and opportunities of two routes in the scope and who are interested to develop a safe, economically viable and circular battery value chain while creating and maintaining a better access to critical raw materials by developing a scalable battery recycling facility.

1.4 Report structure

Following this introductory chapter where the background and motivation for battery recycling in the Netherlands are explained, the report is outlined as follows:

- Chapter 2 elaborates the Blueprint concept,
- Chapter 3 explains the Construction phase of the blueprint concept, defining the battery limits for two chosen process routes and input data used,
- Chapter 4 explains the Utilization phase of the blueprint concept which focuses on process flow diagrams, material, and energy balance and eventually on a preliminary techno-economic assessment,
- Chapter 5 explains the Sensitivity analysis for the most impactful process parameters.
- Finally, Chapters 6 and 7 focus on Conclusions and Recommendations respectively.

Publicly available data, such as patents and academic articles have been the basis for the preliminary techno-economic assessment. However, as the level of details in the data is rather limited, assumptions have been made and can be found in the Appendices or in grey text blocks.

2. The blueprint concept

The idea behind this assignment was to develop a perspective for a recycling line on a large scale and establish a pathway for sustainable recycling of batteries from electric vehicles. Many patents and reports are claiming that different routes give promising results. However, there is a lack of understanding of the process mechanisms as the provided details appear non-transparent. The objective of this task was to build a perspective on a roadmap towards a scalable and flexible recycling process. For this purpose, the publicly available information is used to construct a blueprint for battery recycling and evaluate it through the techno-economic assessment.

The blueprints are not intended to be exhaustive nor to be an accurate description of process units' operations as this would be highly challenging with the available data, but to give a comprehensive representation of an industrial process in terms of its typical availability and main energy and material flows, and services, without disclosing any confidential information. Building the blueprint consists of two phases – construction phase and utilization phase, which are presented in Figure 2 together with the aims.

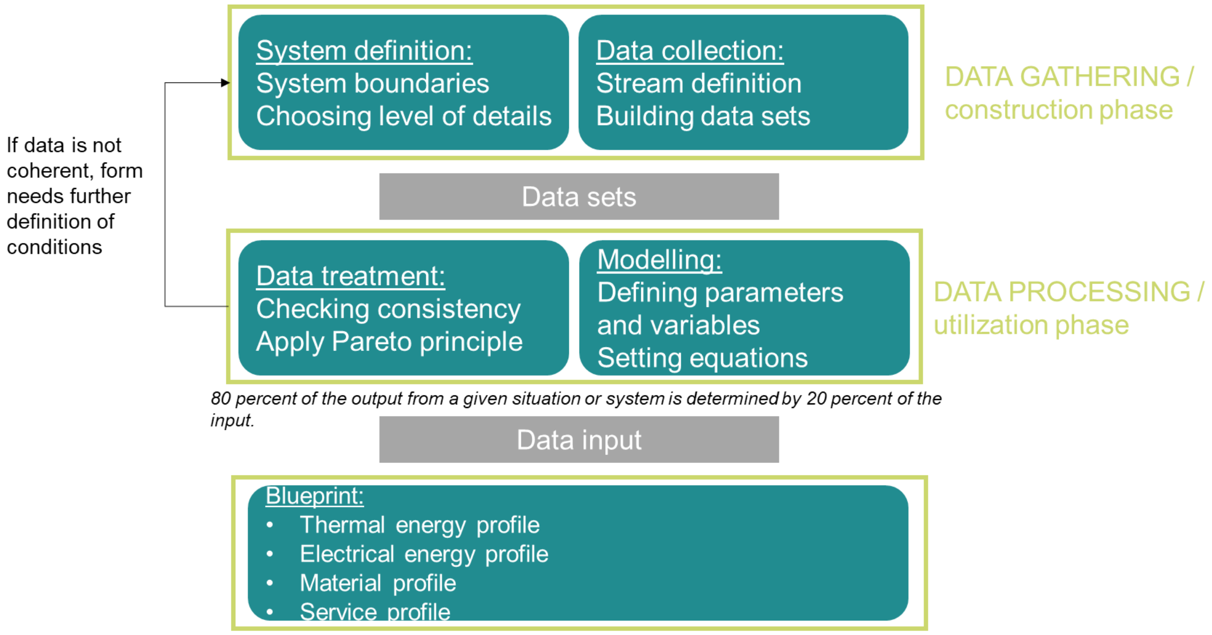


Figure 2: *The Blueprint workflow*

3. Construction phase

The construction phase of this report consisted of the system definition, data collection, and setting the assumptions. The novel processes in this case, which are still not explored enough, have many unknowns. A typical approach in that case is for the unknowns to set different assumptions, which are usually based either on “rules of thumb”, experience, or on the available information for similar types of processes. System definition is made by setting the battery limits, followed by data collection for the selected part of the process. Battery limit represents a defined boundary of responsibilities and clearly indicates what is in the scope of the project or analysis and what is not. When there is no data available, assumptions are made, or sensitivity analysis is done to find a range of parameters under which a certain situation makes sense.

3.1 Battery limits

Two routes are examined in this report: mixed precipitation and solvent extraction. Both routes exclude the pre-treatment step and start from the black mass leaching step, followed by a removal of **non-cathode material** (products of secondary interest, or products with lower market value – graphite, copper, aluminium, and iron) and the **recovery of cathode active material (CAM)** (critical materials as products of primary interest, or products with higher market value – nickel, cobalt, manganese, and lithium). The difference between the two routes lies in the recovery of cathode active material, while the removal of non-cathode material is assumed to be the same.

Wastewater treatment facility falls outside the battery limits, as well as the solvent recovery. For the sake of estimations, assumptions are made to cover the solvent loss and water use. In case of the mixed precipitation route, further treatment of the NMC cake is also outside the battery limits. The outcome is shown in Figure 3.

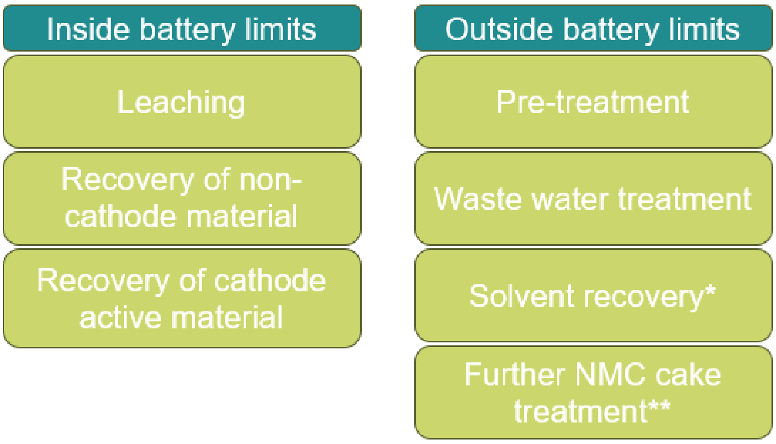


Figure 3: *Definition of battery limit*[1]

---

*Sulfuric acid recovery from the mixed precipitation route is outside of battery limits, as well as sulfuric acid recovery and organic solvents recovery for the solvent extraction route.
**It is assumed process will stop at obtaining NMC hydroxide cake and Lithium carbonate. Any further separation of the NMC cake is not considered in the mixed precipitation route.

3.2 Input data

Once defined what is within the battery limits, block schemes of two processes are made and inlet and outlet streams are mapped. They were the initial guideline for further and more detailed data collection. The main input is black mass, for which two sets of data are investigated. Other consumables use is assumed based on available literature and different patents. Final products are also different for the two routes. In case of mixed precipitation, the final products are lithium carbonate and NMC hydroxide cake, while for the solvent extraction, the final products are lithium carbonate and magnesium-, nickel- and cobalt sulphate, all separated in individual steps. An overview of both process schemes can be made by viewing the block diagram of the two processes in Figure 4 and Figure 5.

3.2.1 Block schemes of processes in scope

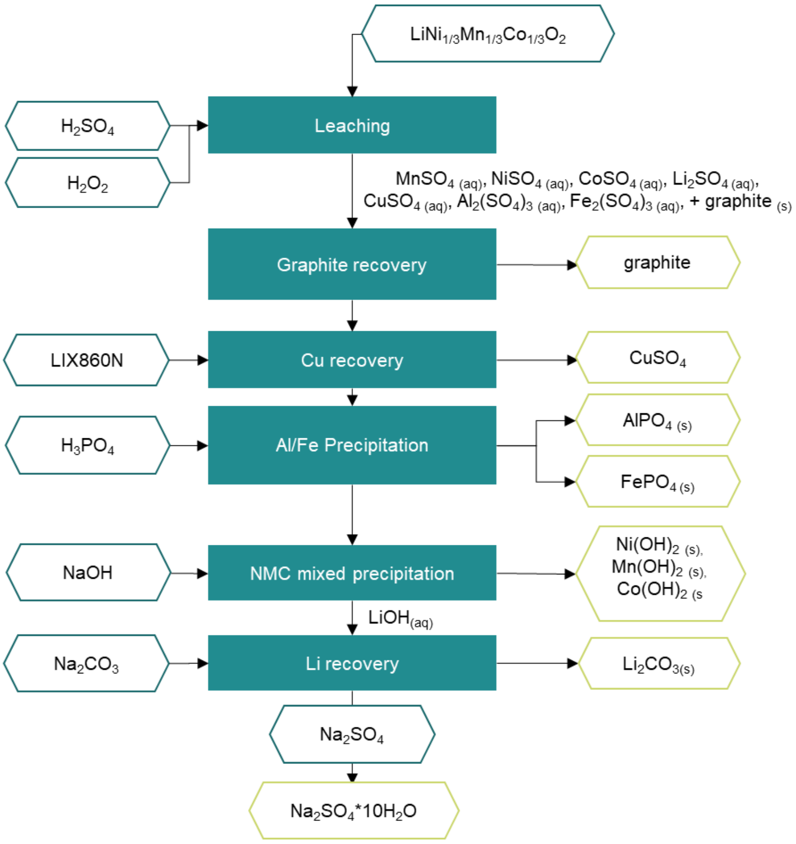


Figure 4: *Block scheme of mixed precipitation route*

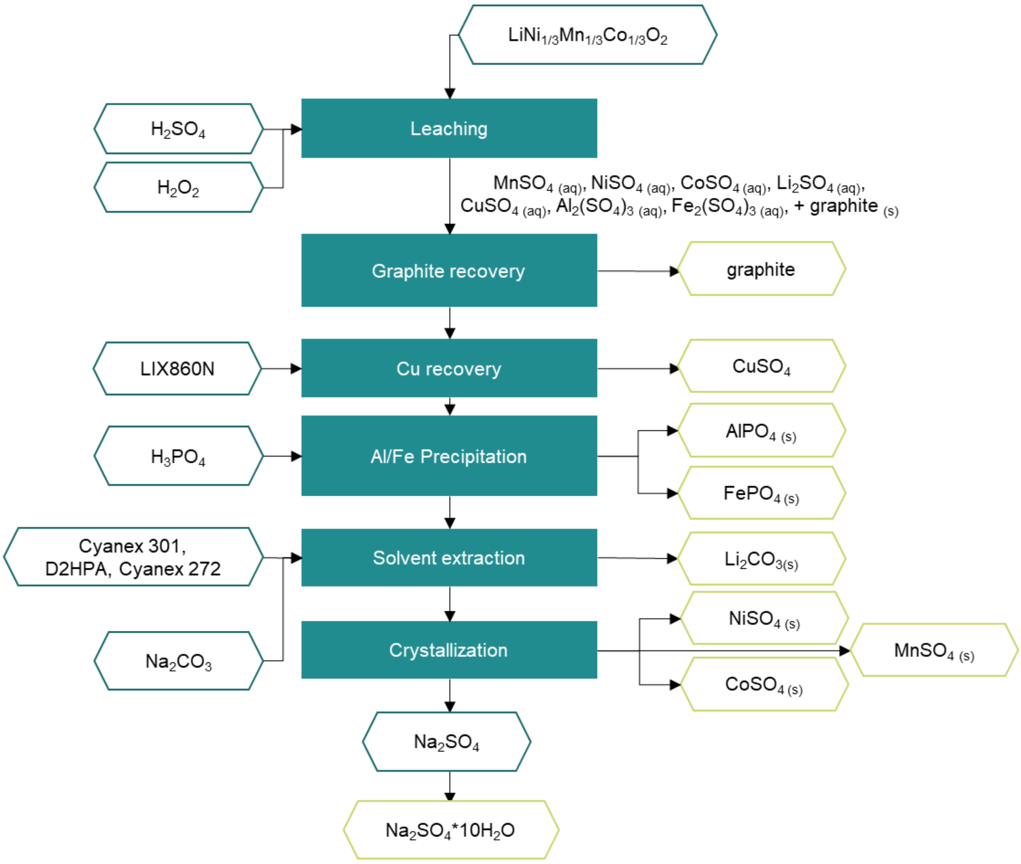


Figure 5: *Block scheme of the solvent extraction route*

3.2.2 Black mass composition

The black mass composition is not uniformly presented in all reports. While some specifically report the black mass composition, others like Diekmann report the composition of the whole battery system consisting of anode and cathode material, cell housing, module periphery, battery system periphery and other. Black mass composition for this work is taken from the Northvolt patent for NMC batteries, while for the purpose of the sensitivity analysis, another set of data is used from Diekmann et al. 2017. The black mass composition used for the study is presented in Table 1. The black mass composition from the table is assumed to present the anode and cathode active material composition including the electrode foils, while the *Other* is assumed to be the battery casing, as on average, 25 to 40 percent of the total battery weight is consisting of metal casing, cables and thermal and battery management systems (EV BOX, 2023).

Table 1: *Composition of black mass (wt.%) used for the study*

	Northvolt NMC rich (Patent No. WO 2020/212363 A1, 2020)	**Diekmann et al. (2017)** (Diekmann, 2017)
Al	3	5.5
Co	9.4	3.1
Ni	9.6	3.1
Mn	9.1	2.8
Li	**4.0**	1.0
Fe	0.1	0.0
Cu	4.3	9.2
Graphite	29.1	8.2
Oxygen	-	4.8
Other (excluded)	31.4	62.3
TOTAL	**100.0**	**100.0**

3.2.3 Solvent use

In the leaching step, black mass is treated with a 2M sulfuric acid. The sulfuric acid is cheap and easy to obtain. It is added to increase the leaching speed and dissolution rate. However, since the leaching efficiency of the sulfuric acid is relatively low, hydrogen peroxide is added as a reducing agent to the sulfuric acid solution during actual operation.

Table 2: *Assumptions for solvent use* (Patent No. WO 2020/212363 A1, 2020), (Patent No. EP 3 535 803 B1, 2018) *(O is organic and A is aqueous phase)*

Step	**Solvent use**
Leaching step (both routes)	2M sulfuric acid in stoichiometric amount with an excess of 10% 30 g/l hydrogen peroxide solution in stoichiometric amount
Copper removal from the pregnant leach solution (both routes)	Mixture of 30 wt.% of LIX860N and 70 wt.% kerosene, O:A ratio = 1:1
Precipitation of aluminium and iron (both routes)	Stoichiometric amount of 85 wt.% phosphoric acid
Separation of manganese and lithium from cobalt and nickel (solvent extraction route)	Mixture of 30 wt.% solution of Cyanex 301 and 70 wt.% kerosene is used, O:A ratio = 1:1
Separation of lithium from manganese (solvent extraction route)	Mixture of 30 wt.% solution D2HP and 70 wt.% kerosene solution is used, O:A ratio = 1:1
Separation of cobalt and nickel (solvent extraction route)	Mixture of 30 wt.% Cyanex 272 and 70 wt.% kerosene is used, O:A ratio = 1:1
Back extraction (solvent extraction route)	2M sulfuric acid is used for stripping the organic solvent, O:A ratio = 1:1

Solvent concentration in the leaching process step, as well as in the extraction and back extraction step, are a point of discussion in the sensitivity analysis, as they directly affect the volume of the stream, sizing of the equipment, as well as the energy requirement and overall costs of the two processes.

3.2.4 Process design

Both routes examined in this report start from the leaching step. Leaching serves to convert the metal oxides in the pretreated cathode materials into ions in the solution to facilitate the subsequent separation and recycling process. Acids used for leaching can be diverse including both inorganic acid (HCl, H_2SO_4, HNO_3, H_3PO_4, etc.), and organic acids like oxalic acid, citric acid, etc. (Aggie Hu, 2024) Valuable metals originating from the cathode end up in a pregnant leach solution, from where metals are recovered in several separation steps like precipitation, solvent extraction, electrowinning, crystallization.

Certain steps are found to lack details, especially specifics for the solvent extraction route: the solvent extraction itself, back extraction and solvent use, the pre-concentration step for the crystallization, crystallization of nickel-, manganese- and cobalt-sulfates in their stable, hydrated forms, inverse solubility of manganese-sulphate, data on the metal salt solubility in ternary mixtures, etc. All this increased the number of assumptions, limiting the reliability of the model.

4. Utilization phase

Process design for the black mass treatment is defined based on the Deliverable 6.2 of the GTDE project (M. B. Rossi, 2024), and represents the basis for the further techno-economic assessment. A generic use case is necessary to perform a comparative analysis of different systems through the sensitivity analysis. The techno-economic analysis is performed for two battery recycling routes. Process designs exclude the pre-treatment in terms of battery dismantling and comminution, but do include final treatment process steps as black mass leaching and downstream processing, which are divided based on removal of non-cathode materials like graphite, copper, aluminium, and iron and on recovery of cathode active material (CAM) like lithium, nickel, cobalt, and manganese, which have a relatively high market value.

A mathematical model is developed for the techno-economic assessment. The model simulates input and output flows, utilities, and their costs, followed by the sizing of equipment needed for the process and accompanying costs for each process.

The sensitivity analysis is performed to understand the sensitivity of the cost segments affecting the overall production price. Sensitivity analysis is performed for selected process and economic parameters by using a model for calculating the production price combined with a What-if analysis.

4.1 Techno-economic analysis

The objective of the techno-economic analysis is to estimate the required investments and the expected profitability of the proposed recycling technology. For the purpose of blueprint development, two hydrometallurgical routes for recovery of the cathode active material are chosen: the mixed precipitation route and the solvent extraction route. Both investigated routes include the leaching step, removal of non-cathode material and recovery of cathode active material and they are the basis for the techno-economic assessment in this report.

After the scope is defined, process flow diagrams are developed and presented in Figure 6 and Figure 7. The next step was to set up the material and energy balances which, serve as the starting point for the technical and economic analysis. Due to the limited design details and the precision of the cost functions, the results of the economic analysis are Class 5 estimates (i.e. the error margin of these estimations is ±30-50%) (Christensen, 2011).

4.2 Process flow diagram

Hydrometallurgical methods generally start with leaching of the target materials: cobalt, nickel, manganese, and lithium in the case of battery recycling. Most industrially scaled processes, if not all, use a solution of H_2SO_4. Optionally, H_2O_2, is used as a reductant (M. B. Rossi, 2024). Directly after the leaching of black mass, the focus moves to the **removal of non-cathode active materials,** which starts with the removal of unreacted solids. This is envisioned to be done by filtration, making use of a filter press or bed filter. The resulting solid can be purified by a two-step flotation process, to isolate a graphite product. Subsequently, after separation, graphite is rinsed with hot water, and sent to the drier after which the final product is obtained. The final product of this step is graphite, which is assumed to be a high purity product which can be re-used in the battery production process. Next step in the process is copper removal, which is envisioned to be done by solvent extraction (with a mixture of LIX860N and kerosene). Copper is then stripped using H_2SO_4 solution and sent to an electrowinning plant to produce a metallic copper cathode material. Iron and aluminium are removed in the next step by precipitation from the leachate by addition of 85 wt.% H_3PO_4 resulting in a mixed phosphate cake and further separated by filtration. Precipitate is washed with water prior to drying. When all other elements contained in batteries are removed, the focus moves to the **recovery of cathode active material (CAM)**. This is where the two different routes are considered as shown in Figure 6 and Figure 7.

Mixed precipitation route. Recovery of CAMs by mixed precipitation is a method designed to recover the materials for direct use as new CAMs. In this way, recycling processes are closely related to the production processes of new batteries (M. B. Rossi, 2024). Before using it in the new battery production process, it is quite possible that the nickel, cobalt, and manganese (NMC)-ratio has to be adjusted by adding some of the virgin materials. The process covers the NMC precipitation by using 50 wt.% NaOH solution, resulting in a cake consisting of NMC salts in the form of hydroxides. The NMC hydroxides are then filtrated from the aqueous phase and dried. Once the cobalt, manganese, and nickel are recovered as their corresponding hydroxides, lithium is recovered separately by precipitation with Na_2CO_3, and is then removed by filtration and dried as Li_2CO_3.

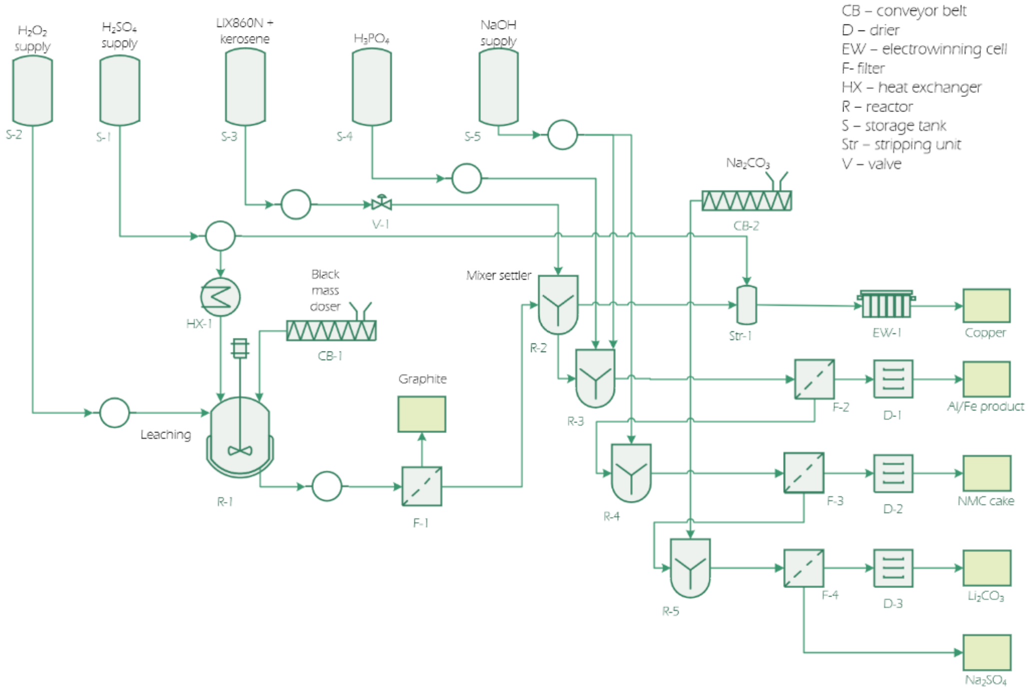


Figure 6: *Schematic process flow diagram of the mixed precipitation route*

Solvent extraction route. This route consists of the removal of metals dissolved in aqueous media (pregnant leach solution), using selective ligands that give rise to soluble complexes in organic solvents. After extraction of the metal complexes into the organic phase, a stripping step is followed, making use of a H_2SO_4 solution to bring the metal ions back to aqueous media. Further on, precipitation and or crystallization is followed by filtration and drying of the target metal allowing the isolation of a metal salt or metallic metal, respectively (M. B. Rossi, 2024).

Cobalt and nickel are extracted from the leachate using Cyanex 301 (bis (2,4,4-trimethylpentyl)-phosphinodithioic acid), dissolved in kerosene. Subsequently, both metals are stripped by means of a H_2SO_4 solution. Following further separation of cobalt using Cyanex 272 (also in kerosene), both metals are crystallized separately, as their corresponding sulphate salts. The raffinate of the nickel and cobalt extraction is enriched in manganese, which is also removed by solvent extraction, making use of DE2HPA (Di(2-ethylhexyl) phosphoric acid) dissolved in kerosene (M. B. Rossi, 2024).

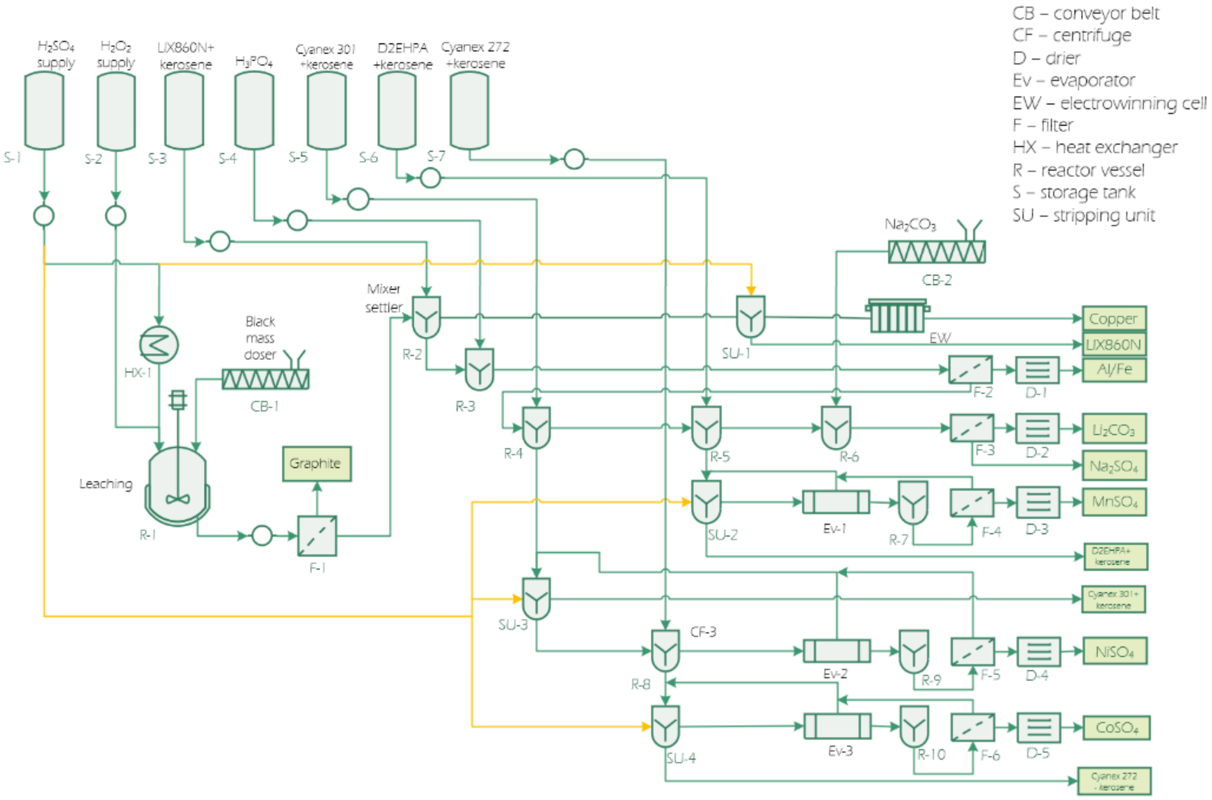


***Figure 7:** Schematic process flow diagram of the solvent extraction route*

The recovery of Li as Li_2CO_3, is usually performed as the last step of these recycling processes. Lithium is concentrated in the filtrate, once all the other metals from the cathode active material have been recovered, by precipitation using Na_2CO_3 alongside sodium sulphate decahydrate ($Na_2SO_4 \cdot 10\ H_2O$). The resulting solid Li_2CO_3 is filtrated, rinsed with hot water, and then dried (M. B. Rossi, 2024).

4.3 Material and energy balance

Based on stoichiometry and assumptions, material balances have been calculated for the proposed commercial scale plants on a basis of 6 kilotonnes per year of black mass feed. Material balances are presented for the base case for mixed precipitation route and for solvent extraction route in Table 3 and Table 4.

The table represents inlet streams in terms of feedstock (black mass) and consumables, and outlet streams in terms of products, streams in loop (consumables with certain losses that will be recycled back in the system with an addition of a virgin materials) and waste streams which would be treated externally.

All solvents (LIX860N, Cyanex 301, Cyanex 272, Kerosene, H_2SO_4) must be reclaimed at certain point, but this has not been considered in this work. Process losses are estimated based on the assumed individual process efficiencies of each step. Process losses are estimated as the difference of inlet and outlet stream and they are assumed to be covered by adding virgin material. Waste processing costs are not assumed within this work.

Table 5 presents a summary of plant energy requirements for mixed precipitation and solvent extraction routes. The energy needs are split into three categories: heat, cooling, and electricity. Electricity needs consist of pumping, mixing, filtration, electrowinning and transporting of solid feeds with conveyor belt.

Table 3: *Material balance for mixed precipitation route on a basis of 1 kg black mass feed*

Component	IN		OUT	Rest
	kg feed	kg/kg feed	kg/kg feed	kg/kg feed
Black mass	1			
H_2SO_4		7.64		
H_2O_2		4.41		
NaOH		18.24		
H_3PO_4		0.17		
Na_2CO_3		0.64		
LIX860N		12.71	12.70	
Graphite			0.34	
$CuSO_4$			0.13	
NMC-OH cake			0.53	
Li_2CO_3			0.21	
Na_2SO_4			3.85	
$AlPO_4/FePO_4$			0.16	
Waste water				26.66
Oxygen				0.06
CO_2				0.14
loss				0.01

Table 4: *Material balance for solvent extraction route calculated on a basis of 1 kg black mass feed*

Component	IN		OUT	Rest
	kg feed	kg/kg feed	kg/kg feed	kg/kg feed
Black mass	1			
H_2SO_4		54.04		47.13
H_2O_2		4.41		
H_3PO_4		0.17		
Na_2CO_3		2.12		
LIX860N		3.46	3.46	
D2HPA		3.22	3.19	
Cyanex 301		3.39	3.37	
Cyanex 272		3.77	3.74	
Kerosene		32.28	3.11	
Graphite			0.34	
$CuSO_4$			0.14	
$MnSO_4$			0.29	
$CoSO_4$			0.29	
$NiSO_4$			0.30	
Li_2CO_3			0.21	
$Na_2SO_4*10H_2O$			1.94	
$AlPO_4$			0.16	
$FePO_4$			0.00	
Waste water				10.92
off-gas				0.06
CO_2				0.19
losses				0.00

Table 5: *Plant energy needs*

	Mixed precipitation	**Solvent extraction**
Type	Power [kW/kg feed]	
Heating	3.52	23.17
Cooling	2.15	22.55
Electricity	0.16	1.15

The big difference in plant energy needs for two routes can be explained with the energy intensity of the solvent extraction route and specifically evaporative crystallization step.

4.4 Equipment sizing and costing

Equipment costs are based on equipment sizing and they also take into consideration and are multiplied by factors like location of the plant (Netherlands) and the material of construction. It is assumed that the material is corrosion resistant, and it has the same price as stainless steel SS304. The starting point assumptions for the two processes are summarized in Table 6.

Table 6: *Overview of key techno-economic assessment assumptions*

Scale: 6 kta of **black mass feed**
Location: Netherlands (location factor 1.19)
Material of equipment construction: Similar to SS304 (material factor 1.3)
Year for costs and value: 2023 (CEPCI of 797.9)

A novelty factor of 10% is added to the equipment cost (for a new and unproven process, new process, redesigned, licensed or proven process) resulting in total equipment cost (TEC). Finally, a factor of 10% is added to the total equipment cost for the equipment delivery, resulting in the total delivered cost as shown in Table 19.

4.5 Economic investment

4.5.1 Capital investment

The total direct cost of two routes is estimated as a sum of cost components listed in Table 7 while their ratio is shown in Figure 8. The individual contribution of each cost component is calculated by multiplying the Total Delivered Cost with a specific factor for the given cost component. Factors may differ depending on the type of plant – is it a solid processing, solid-fluid processing or fluid processing plant. The total direct costs for two routes are presented in Figure 9 while the factors applied in estimating total direct costs can be found in the Appendix B, Table 19.

Table 7: *Cost components of the Total Direct Cost (TDC)*

Total direct cost	Total delivered cost Purchased equipment installation Instrumentation & Controls (installed) Piping (installed) Electrical systems (installed) Buildings (including services) Yard improvements Service facilities (installed)

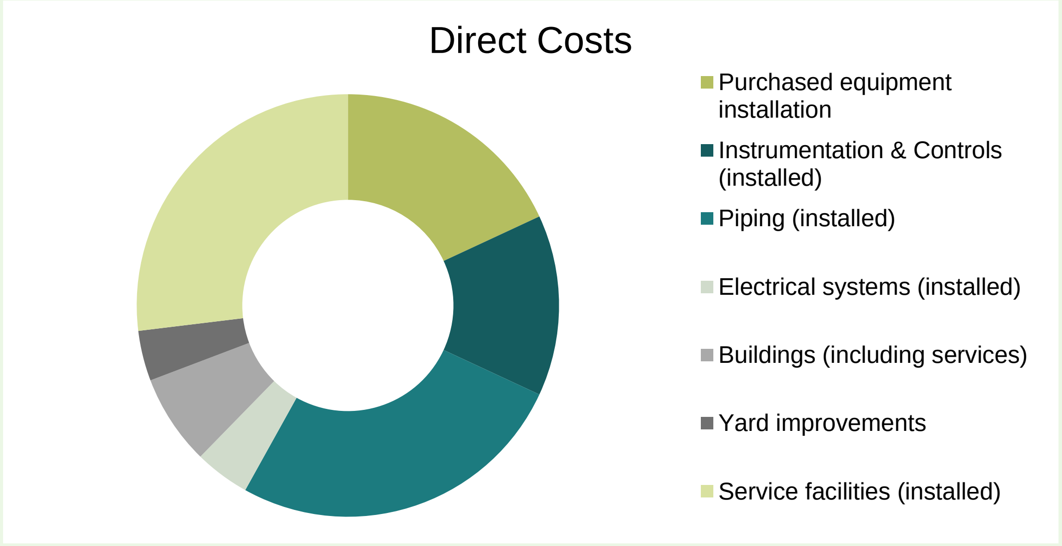


Figure 8: *Ratio of factors in calculating direct costs for a solid-fluid processing costs*

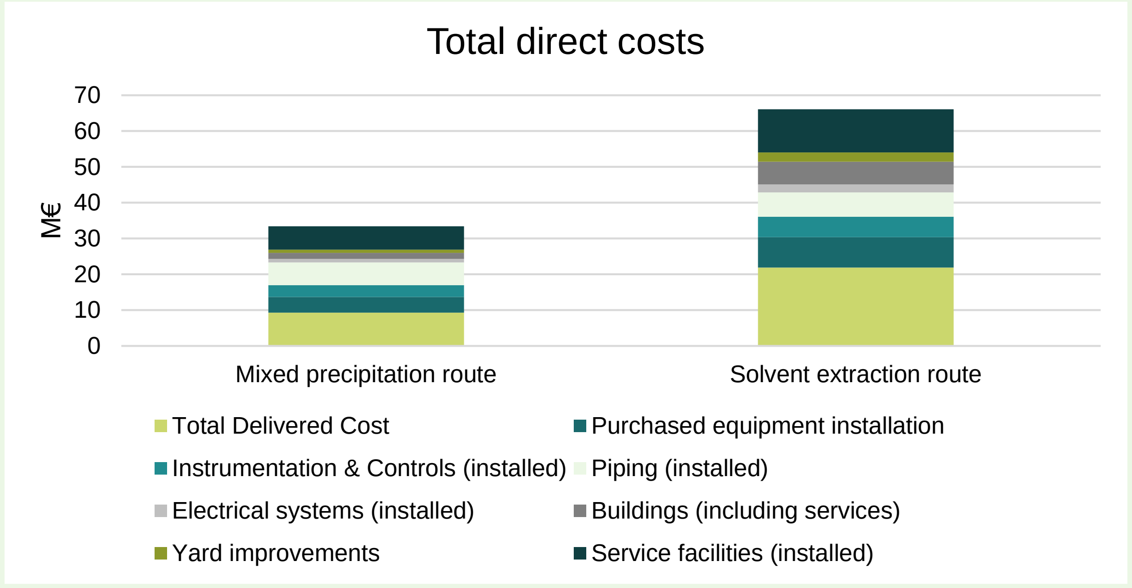


Figure 9: *Comparison of total direct costs for mixed precipitation and solvent recovery route*

Total delivered cost followed by installed service facilities and installation of purchased equipment represent the highest part of the investment.

The difference in total direct costs for mixed precipitation route and solvent extraction route reflects the process complexity. Both estimations cover the leaching step, removal of non-cathode material and recovery of cathode active material. First two parts of the process are assumed to be the same. Therefore, the whole difference shows the complexity of the cathode active material recovery in the solvent extraction route and its costs. The difference originates from having more equipment, from higher equipment size and energy intensity of the process. Solvent extraction route has a downstream process consisting of solvent extraction steps, back extraction steps, evaporative crystallization and filtration next to drying for each metal recovery (manganese, cobalt, nickel and to a great extent lithium). Mixed precipitation route stops at lithium carbonate and NMC hydroxide cake precipitation, without further separation of metals.

The Total Indirect Cost (TIC) is estimated as a sum of cost components listed in Table 8. Their ratio applied to indirect cost components for solid-liquid processing plant are shown in Figure 10 and the total indirect costs for two routes are presented in Figure 11. Factors applied in estimating total indirect costs can be found in the Appendix B, Table 20.

Table 8: *Cost components of Total Indirect Costs*

Total Indirect Costs	Engineering and supervision Construction expenses Legal expenses Contractor's fee Contingency

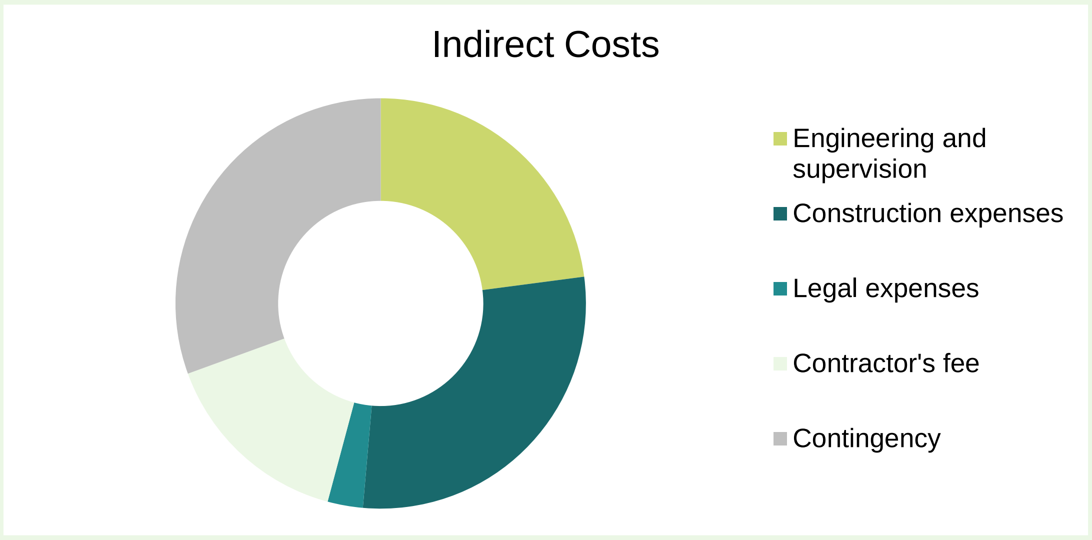


Figure 10: *Ratio of factors in calculating indirect costs for a solid-fluid processing costs*

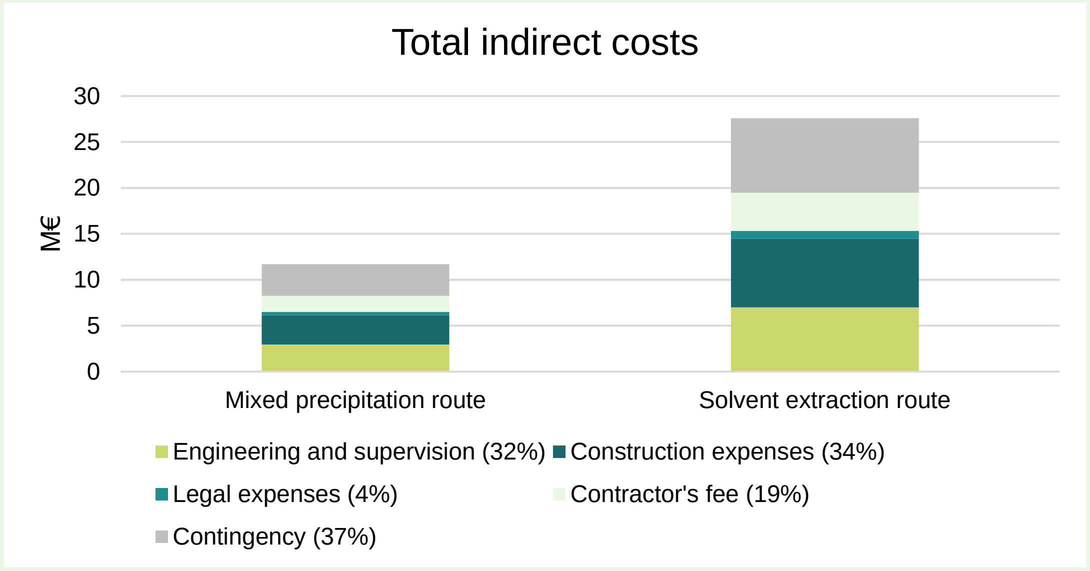


Figure 11: *Comparison of total indirect costs mixed precipitation and solvent recovery route*

Just as in total direct costs, in Figure 11 is shown the impact of additional separation steps in the solvent extraction route on indirect costs. In addition to the total direct and indirect costs, working capital for the plant has to be provided. This is based on 75% of the total equipment delivery cost, bringing the total capital investment (TCI) to M€46.7 in case of mixed precipitation route and M€110.1 in case of solvent extraction route. This is shown in the Figure 13.

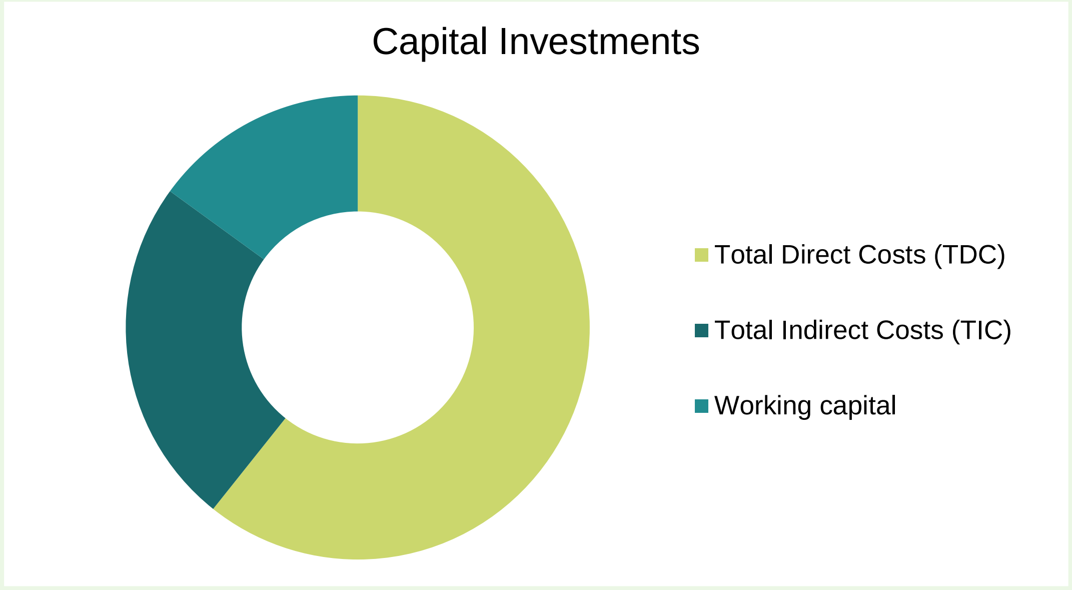


Figure 12: *Ratio of individual components in total capital investments for a solid-fluid processing costs*

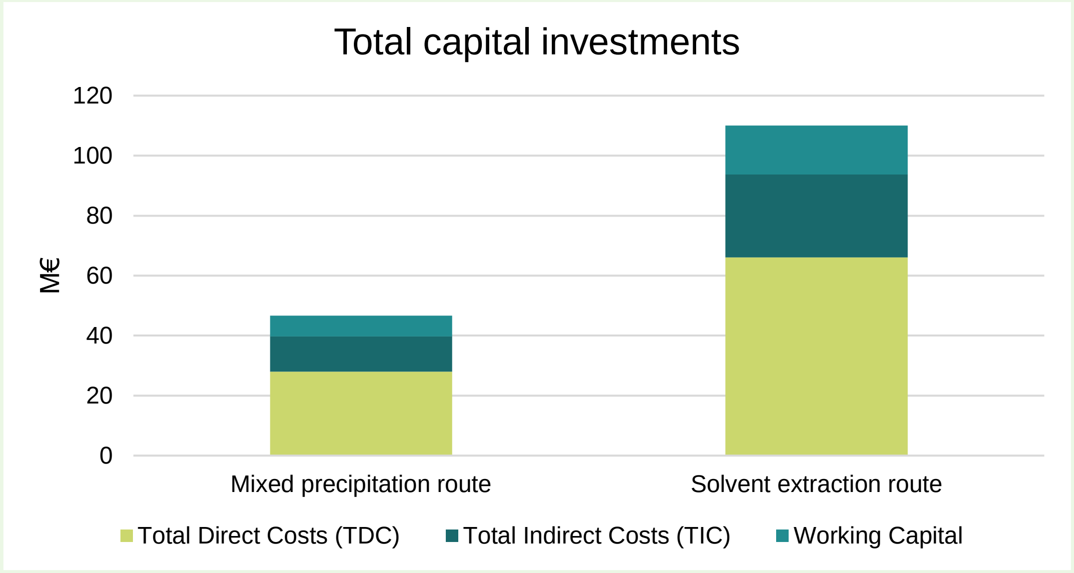


Figure 13: *Total capital investments for mixed precipitation and solvent recovery route*

4.5.2 Production costs

Production costs are composed of material, labor, energy, and other overhead costs. Material costs are presented in Table 21 and Table 22. Labor costs are estimated in the Appendix B, where the main assumptions are:

Number of workers: 2

Number shifts: 3

Number of days per year: 330

Hourly rate: 40 €/hr

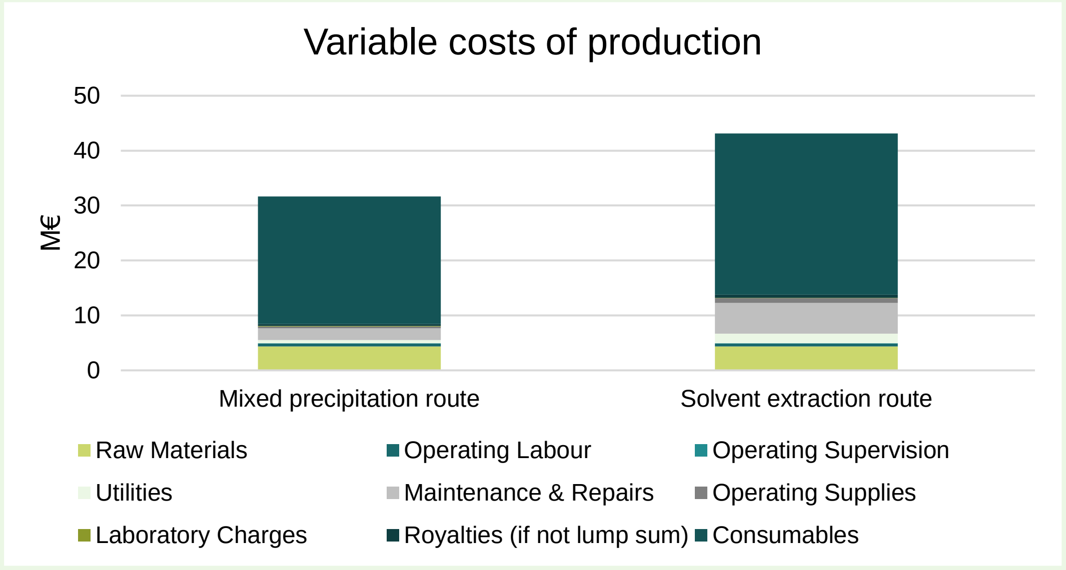


***Figure 14:** Variable costs of production for mixed precipitation and solvent recovery route*

As can be observed in Figure 14, the biggest differences between two processes are in consumables. Solvent extraction process has more chemicals used per kilogram of feed and more losses that have to be compensated with virgin solvents, increasing the cost of consumables.

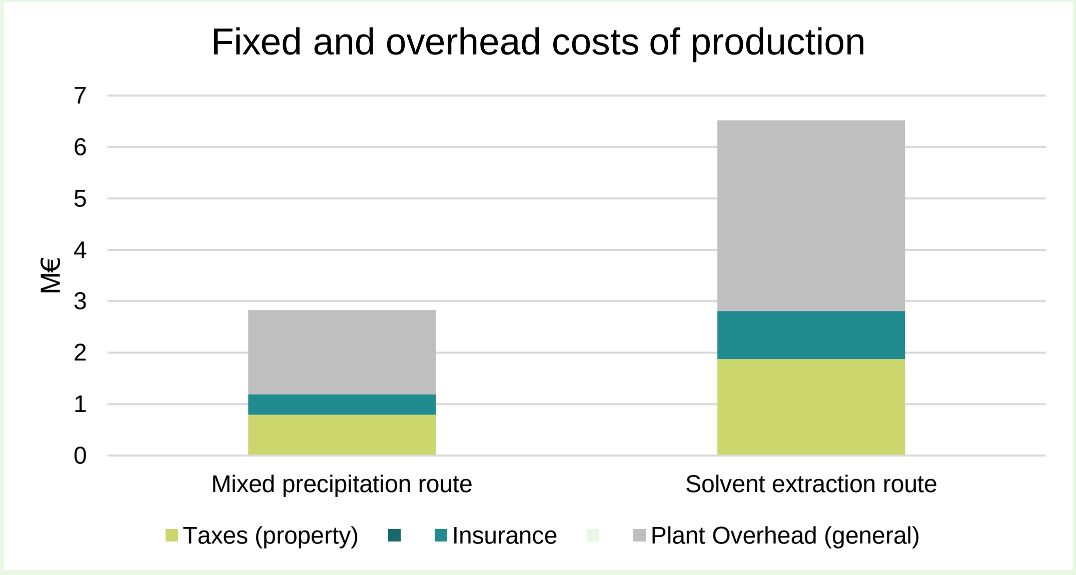


***Figure 15:** Overview of fixed and overhead costs of production*

General costs of production and total costs of production (without depreciation) are presented in Figure 16 and Figure 17.

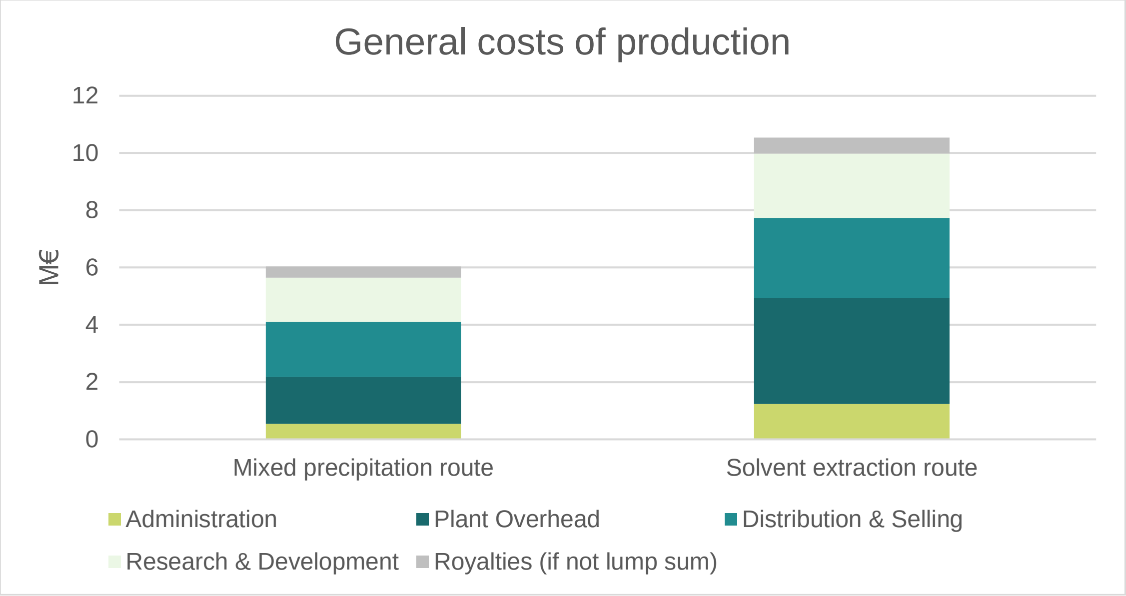


Figure 16: *General production costs for mixed precipitation and solvent recovery route*

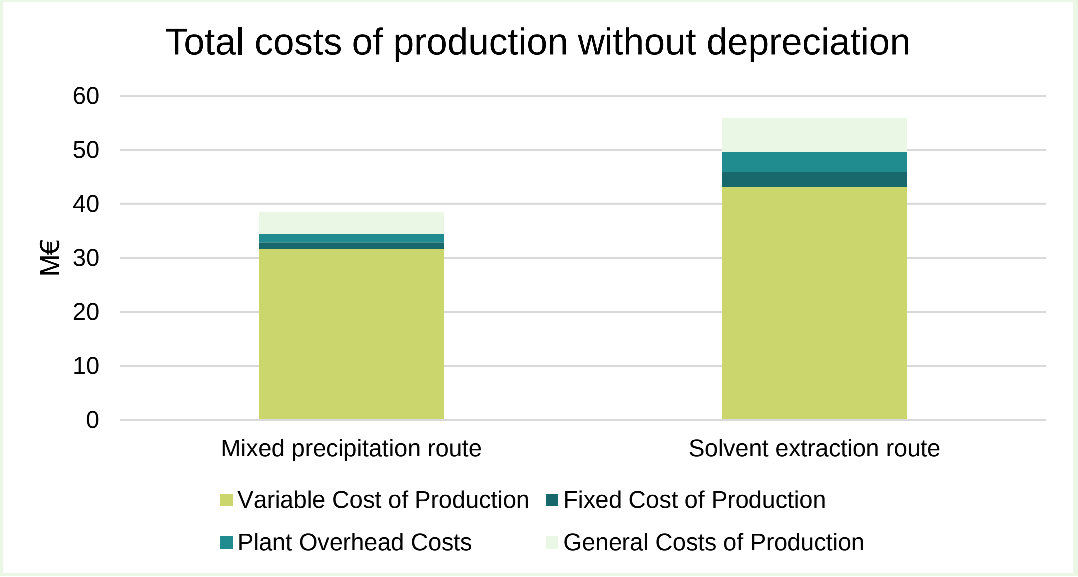


Figure 17: *Total costs of production without depreciation for mixed precipitation and solvent recovery route*

5. Sensitivity analysis

During the material and energy balance execution, it is noticed that the conditions assumed for base cases lead to large volumes that result in higher investment costs. Two routes clearly differ by the quality of the obtained product which significantly affects the costs. Mixed precipitation route is shown to be more economically viable due to lower complexity than the solvent recovery route, but with limited application On the other hand, the solvent extraction route is more complex and more sensitive to different choices (choice of the type of solvent, aqueous to organic ratio, acid concentration, etc.), so the sensitivity analysis is conducted for the solvent extraction route for several process parameters.

5.1 Battery compositions

The battery composition directly affects the potential revenue as the recovered cathode active material has a high value on the market, as metals recovered from it belong to critical raw materials. In two investigated battery compositions, the composition of Diekmann has more non-cathode material, especially copper and less active cathode material. Having less active cathode material means less revenue, but having more non-cathode material like aluminium, iron, and especially copper means higher investment costs. Separation of copper has shown to be the most demanding process step in the removal of non-cathode active material part of the process, first for having the electrowinning cell which is the equipment unit with the highest cost, but also for using the solvents for extraction and back-extraction. Comparison of capital investments and total cost of production for two black mass compositions on the example of solvent extraction route are presented in Figure 18 and Figure 19.

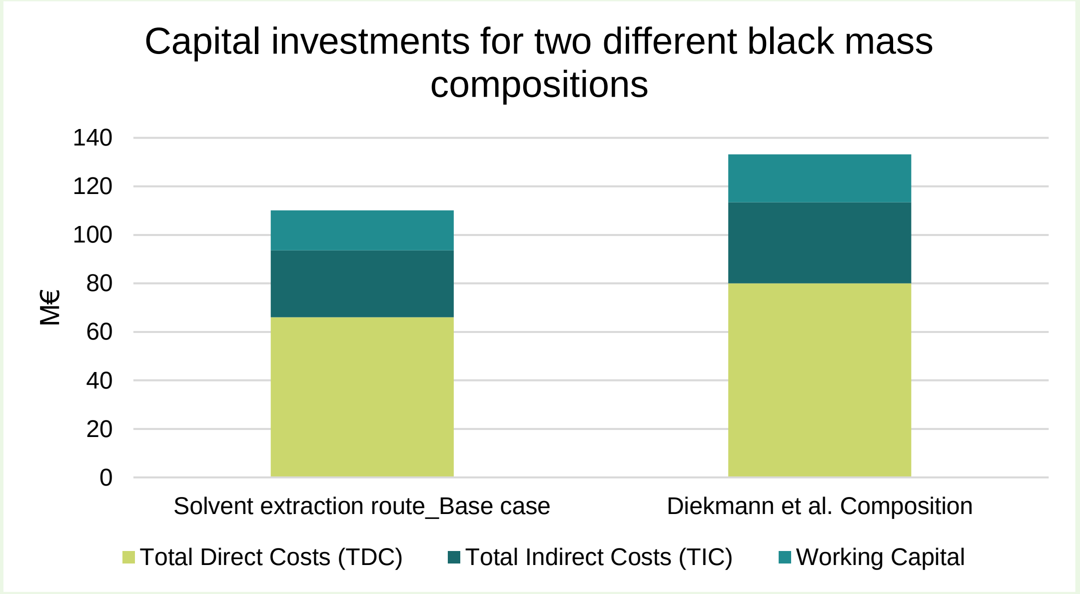


Figure 18: *Comparison of capital investment costs for two compositions of black mass for the solvent extraction route*

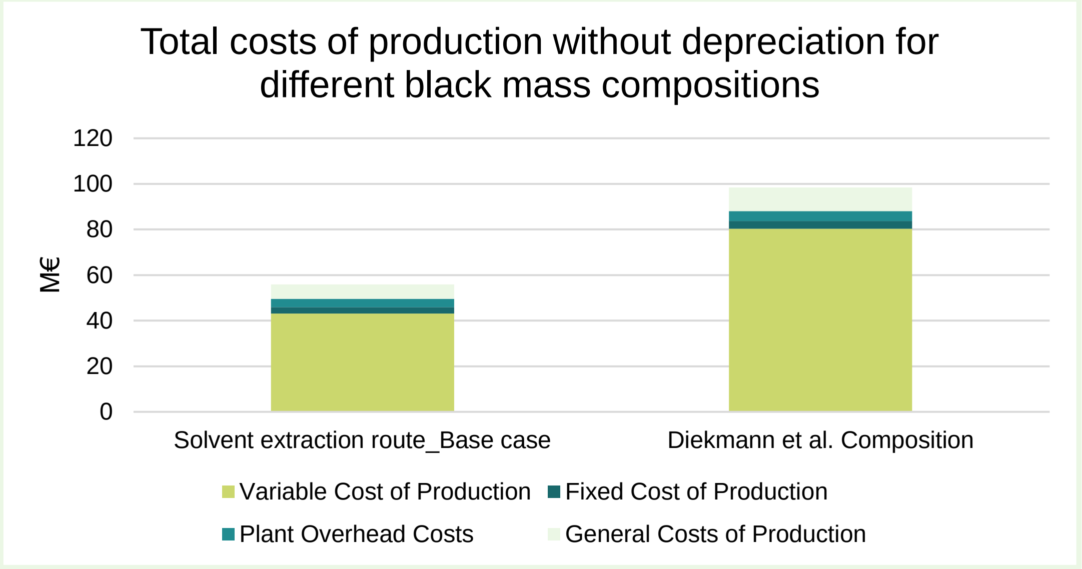


Figure 19: *Comparison of total costs of production (without depreciation) for two compositions of black mass for the solvent extraction route*

5.2 H_2SO_4 concentration

During the execution of material and energy balance, it is observed that large volumes of relatively diluted sulphuric acid are used, resulting in increased capital investment due to equipment sizing and increased operating costs due to the solvent removal in the evaporative crystallization step. The sensitivity analysis is conducted for different concentrations of sulfuric acid and the conclusion is that by increasing the concentration of sulphuric acid, capital investment can decrease for up to 25% in comparison with the base case for the solvent extraction route (2M sulphuric acid). At the same time, the electricity consumption can be reduced by 43.4% in comparison with the base case, and steam requirement can be reduced for up to 78.7% in comparison with the base case, as there is less solvent to be evaporated. Effects of the sulfuric acid concentration on the total capital investment costs and total costs of production for the solvent extraction route are presented in Figure 20 and Figure 21. It has to be noted that the higher sulfuric acid concentration might affect the choice of material used for the equipment asking for a more resistant and expensive one, but this is not taken into account for the purpose of this sensitivity analysis.

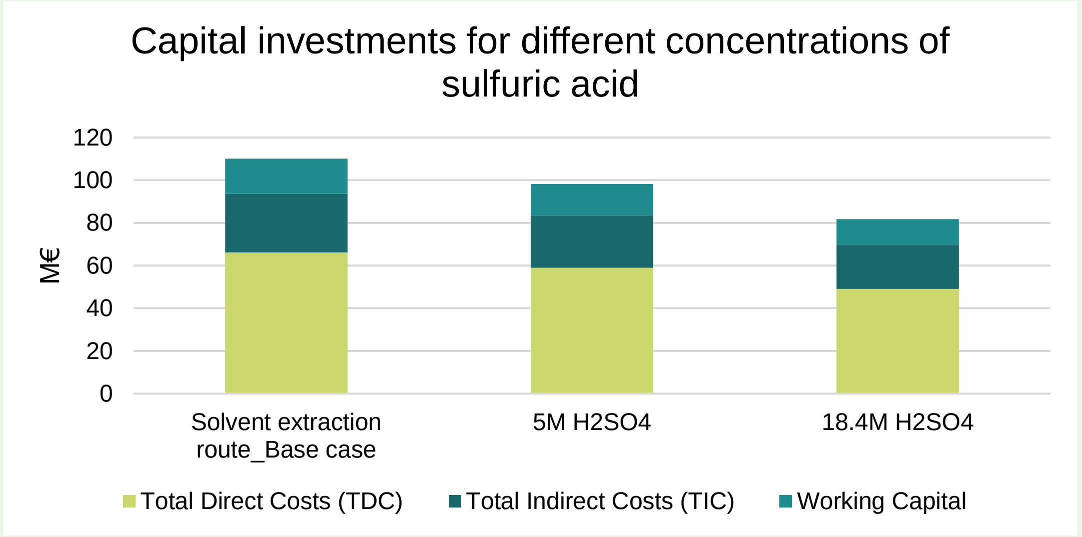


Figure 20: *Comparison of total capital investment costs for 2M sulfuric acid concentration (Base case) with 5M and 18.4 M sulfuric acid concentration for the solvent extraction route*

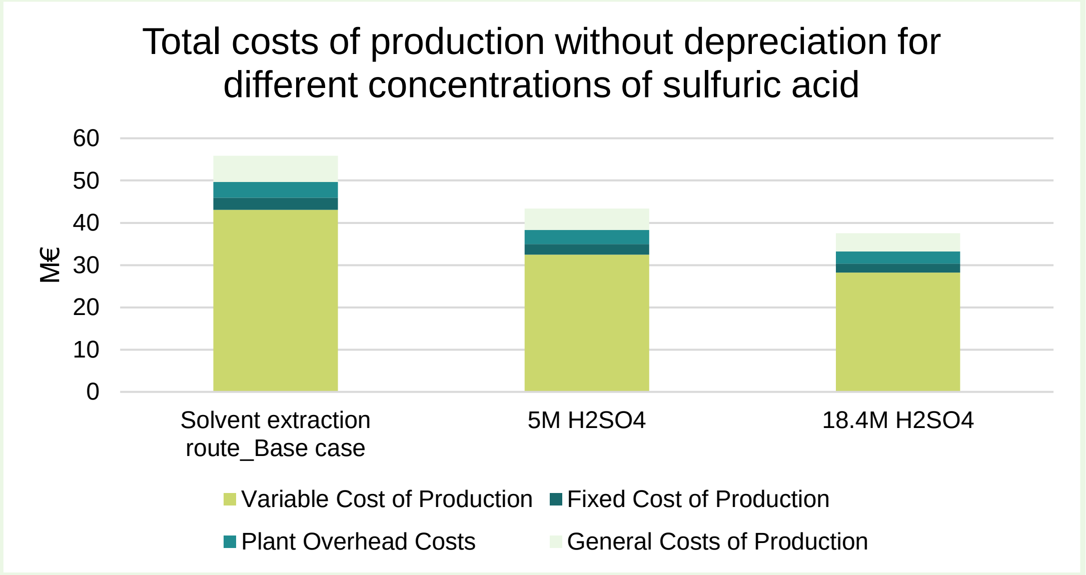


Figure 21: *Comparison of total costs of production (without depreciation) for 2M sulfuric acid concentration (Base case) with 5M and 18.4 M sulfuric acid concentration for the solvent extraction route*

5.3 Aqueous to Organic phase ratio

Following the patents on solvent recovery, it is taken for the base case that the ratio of the aqueous to organic phase is 1:1 in both solvent extraction and back extraction step. However, having already large volumes of 2M sulfuric acid means this will lead to high volumes of organic solvent used in the solvent extraction step and again of 2M sulfuric acid in the back-extraction step. For this purpose, a sensitivity analysis is conducted for 4 scenarios (A to D) where the aqueous to organic phase ratio for solvent extraction is varied from 1:1 to 3:1 in the solvent extraction and 1:1 to 1:3 in the back extraction step as presented in the Table 9.

Table 9: *Sensitivity analysis scenarios for varying the aqueous to organic ratio in solvent extraction and back extraction step as presented in Figure 22 and Figure 23*

Scenario	**A**	**B**	**C**	**D**
Solvent extraction	A:O = 3:1	A:O = 3:1	A:O = 2:1	A:O = 2:1
Back extraction step	A:O = 1:3	A:O = 1:1	A:O = 1:2	A:O = 1:1

The sensitivity analysis shows that it is possible to reduce the total capital investment for approximately 21% if the aqueous to organic phase ratio would be 2:1 in the extraction and 1:2 in the back-extraction step, and around 31.5% % if the aqueous to organic phase ratio would be 3:1 in the extraction and 1:3 in the back-extraction step. This is presented in Figure 22.

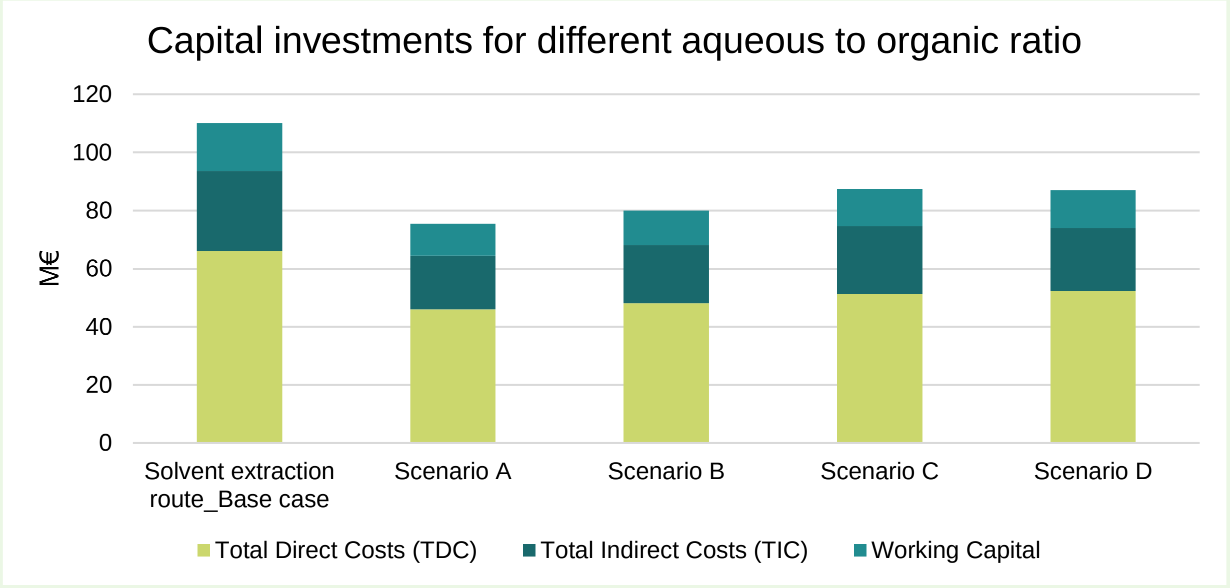


Figure 22: *Comparison of total capital investment costs for solvent extraction route for a process where ratio A:O=1:1 is in the extraction step (E) and A:O=1:1 ratio is in back-extraction step (BE) (Base case) with processes where A:O=3:1 in the extraction step and A:O=1:3 in the back-extraction step (Scenario A), A:O=3:1 in the extraction step and A:O=1:1 in the back-extraction step (Scenario B), A:O=2:1 in the extraction step and A:O=1:2 in the back-extraction step (Scenario C), A:O=2:1 in the extraction step and A:O=1:1 in the back-extraction step (Scenario D)*

In case of the total costs of the production, going to aqueous to organic phase ratio of 2:1 in the extraction and 1:2 in the back-extraction step would result in decrease of total costs of production by 19%, and around 23.7% if the aqueous to organic phase ratio would be 3:1 in the extraction and 1:3 in the back-extraction step. This is presented in Figure 23.

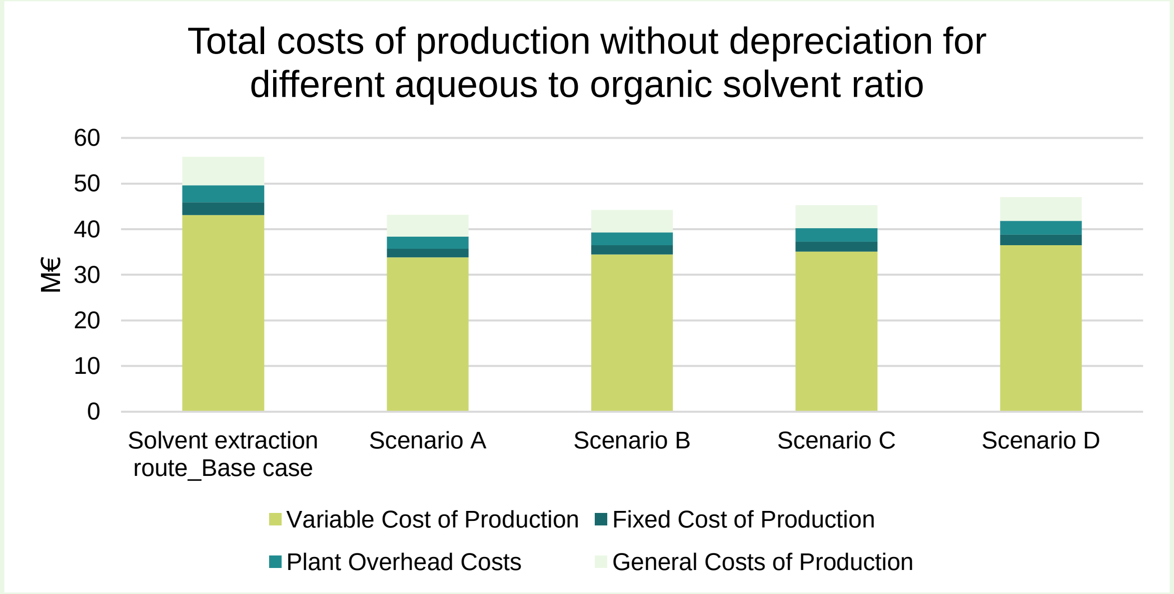


Figure 23: *Comparison of total costs of production (without depreciation) for solvent extraction route for a process where ratio A:O=1:1 is in the extraction step and A:O=1:1 ratio is in back-extraction step (Base case) with processes where A:O=3:1 in the extraction step and A:O=1:3 in the back-extraction step (Scenario A), A:O=3:1 in the extraction step and A:O=1:1 in the back-extraction step (Scenario B), A:O=2:1 in the extraction step and A:O=1:2 in the back-extraction step (Scenario C), A:O=2:1 in the extraction step and A:O=1:1 in the back-extraction step (Scenario D)*

6. Conclusions

Within this assessment, a blueprint for a recycling facility was set-up, to provide an indication of what is needed to establish in the Netherlands for a battery recycling process with aim to recover critical metals from end-of-life batteries and to understand the major challenges and opportunities for such a process. The work on the blueprint consisted of a construction and a utilization phase. The developed blueprint gave a representation of two battery recycling process routes supported with main energy and material flows and services. Based on these two defined process routes, a preliminary techno-economic assessment was performed, and sensitivity analysis was provided for specific process parameters.

The primary goal of this work was to present an understanding of opportunities and challenges of both process routes for recovery of critical metals from end-of-life batteries, which are presented in Table 10. It is noted that it was not intended to compare the two routes in a fully accurate quantitative way or to show exact investments. Additionally, the development of the methodology used in this work was a major objective of this work because this approach can be applied to other critical metal recovery processes or to new iterations of the studied process routes but with better specified and more detailed information, preferably from experimental tests.

Table 10: *Opportunities and challenges for two chosen battery recycling routes*

	Mixed precipitation route	**Solvent extraction route**
Opportunities	• Relatively simple process • Less chemicals used than in other route • Less complex process and hence lower investment costs	• High degree of critical raw material (cathode active material) recovery • Opportunity of higher revenues on final products (separate metal salts) • Possibility to use the final products for applications other than batteries • More flexibility with respect to battery composition as input material
Challenges	• Final product (mixed metal hydroxide cake) less suitable for direct application and hence lower value • Final product must be subjected to further treatment, this results in less environmentally friendly process • Part of the critical raw materials remain “trapped” in a cake • Large waste salt brine stream to be discharged	• More complex process and larger volume streams and hence higher investment cost • Less environmentally friendly process because of larger quantity of chemicals used per kg of black mass treated • Large waste salt brine stream to be discharged • More energy intensive process, especially in pre-concentration of the crystallizer inlet streams

The overview of total investment costs and total costs of production for the mixed precipitation route, base case of solvent extraction route and its sensitivity analysis is given in Figure 24 and Figure 25. Based on the preliminary techno-economic assessment conducted for these two routes, it can be seen that under certain conditions, total investment costs can be decreased for the solvent extraction route, but even the optimum case from the sensitivity analysis has 1.6 times higher total investment costs than the mixed precipitation route. At the same time, total production costs from the optimum case from the sensitivity analysis for the solvent extraction route are 1.1 times higher than the total production costs estimated for the mixed precipitation route.

It must be noted that all calculations were made on the information available in the patents on which this study is based on. The quantity and level of detail of this information, however, was limited, resulting in a large number of assumptions and therefore it adds to lower accuracy of results.

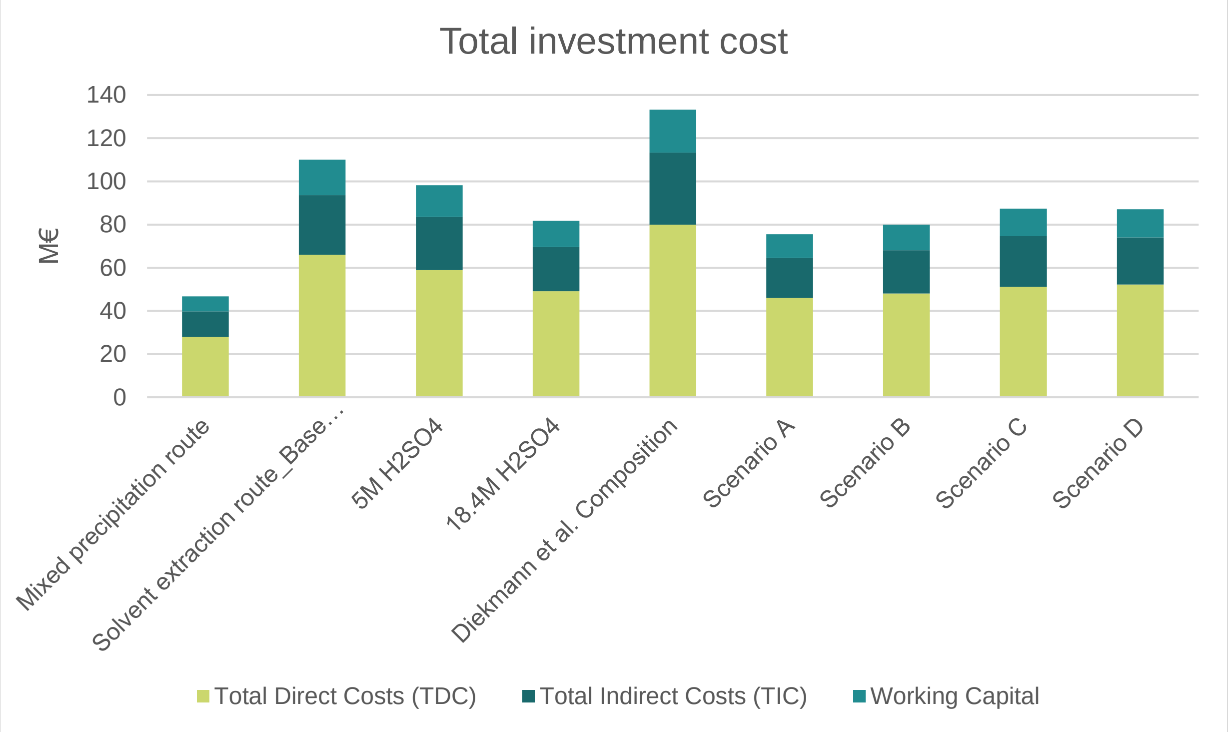


Figure 24: *Summary of results on total investment costs for base case on solvent extraction and different scenarios*

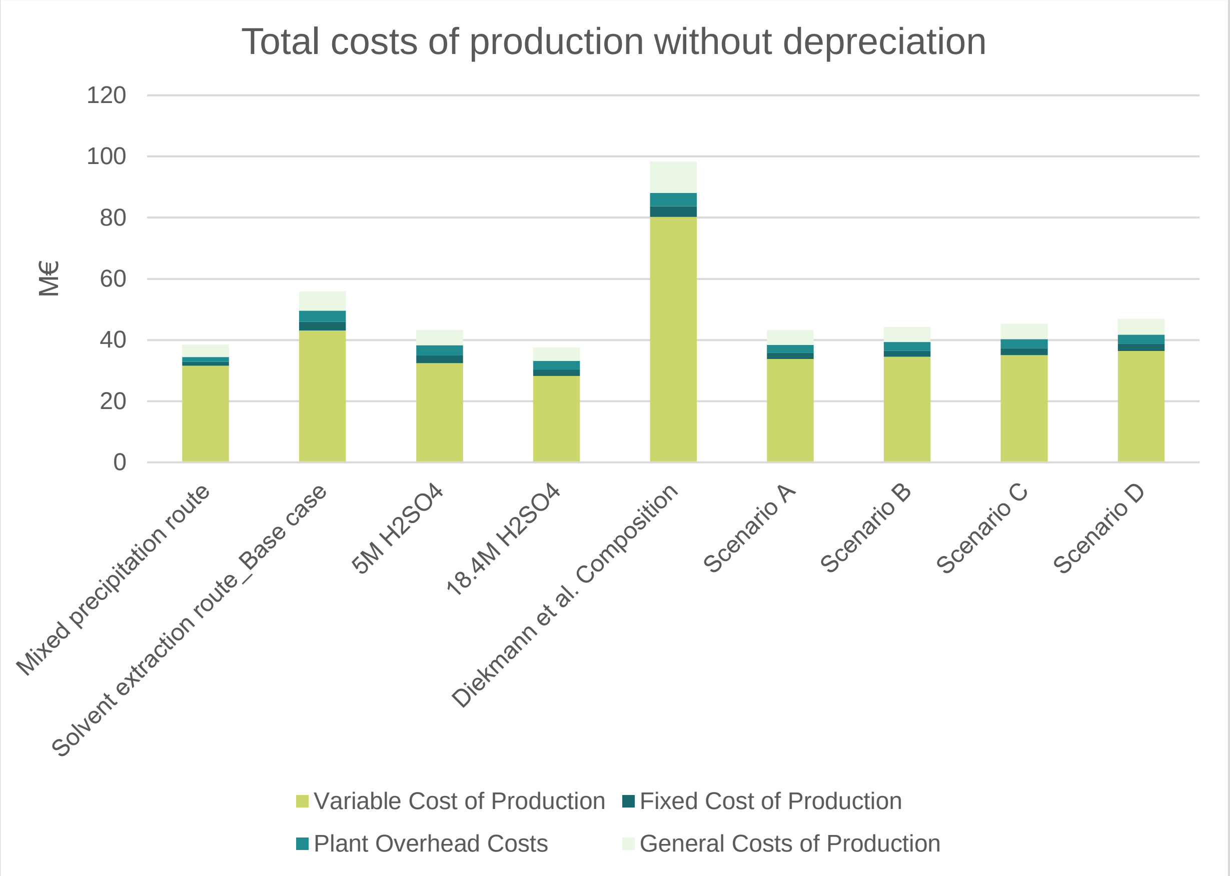


Figure 25: *Summary of results on total costs of production for base case on solvent extraction and different scenarios*

7. Discussion and Recommendations

From the aspects of widely different process complexity and of different final products with corresponding different application possibilities and hence different value, it was not an easy task to compare the two hydrometallurgical processes that were selected for this study. The same challenge was observed in an earlier part of this research as presented in the report *Literature assessment of Li-ion battery recycling steps* (M. B. Rossi, 2024), where more different processes for recovery of critical metals from end-of-life batteries were compared.

Having the NMC cake as the final product, the mixed precipitation route has a limited application in the industry. The most likely option is that such a mixed precipitation route is used as an additional process as part of a battery production facility, so the final product from the recycling process (NMC cake) can be re-used at the same location, probably with correction of the cake composition by adding virgin materials. Otherwise, the NMC cake has to be treated further to separate all the critical materials individually, probably again via the solvent extraction route itself.

During the execution of this task, some gaps were identified which might be of interest to solve in a potential follow-up study. The gaps mostly reflect the missing data and the uncertainty on what the “big picture” would look like. Therefore, a two-way approach is recommended to match the envisioned objective and the practice of working out this vision: top-down and bottom-up analysis of these processes.

The top-down analysis would serve to gain better insights into the Dutch opportunities for battery recycling. It is recommended to orchestrate a consortium that will cover the whole or a segment of a value chain of interest and identify national needs as the outcome, since battery recycling needs strong support with the battery waste collection and pretreatment and further integrating the recovered material into the battery production process. It is recommended to start from the higher level and make a market analysis on potential partners, opportunities, risks, and alternatives. The next step after this assignment would be to build a scenario for a national roll-out of recycling lines and define under which conditions that scenario would have a positive business case. To support that, an improved insight into the future circumstances for battery recycling in the Netherlands suggests that the process scale of 6 kilotonnes of black mass per year, which is used in this report, is on the lower side, while the process economy would benefit from the higher scale.

The bottom-up analysis would follow a similar line as in this report. The technical analysis presented in this work indicates the lack of proper information on solvent use, solvent extraction, pre-concentration, and crystallization, with ranges that could affect the life cycle analysis and question the logic behind the choice of the best possible technology for battery recycling. Therefore, once sufficient, and reliable information is obtained from experiments on different scales, internally or externally through collaboration with partners, it is suggested to make another iteration of the techno-economic assessment which would be further supported by the life cycle assessment. In this way, both types of analysis can benefit from each other.

8. References

Blackridge Research & Consulting. (2024, June 24). *Top 10 Battery Manufacturers in Europe*. Retrieved from www.blackridgeresearch.com: https://www.blackridgeresearch.com/blog/latest-list-of-largest-top-battery-manufacturers-manufacturing-companies-in-europe

Aggie Hu. (2024, May 07). *A brief understanding of spent NMC batteries recycling process*. Retrieved from mysteel.net: https://www.mysteel.net/news/5053575-a-brief-understanding-of-spent-nmc-batteries-recycling-process

Allanore, S. C. (2020). *Estimating the Capital Costs of Electrowinning Processes.* The Electrochemical Society Interface 29, 2.

Christensen, P. a. (2011). *Cost Estimate Classification System – AsApplied in Engineering, Procurement, and Construction for the Process Industries (TCM Framework: 7.3 - Cost Estimating and Budgeting).* Morgantown: AACE International.

Diekmann, J. (2017). Ecological Recycling of Lithium-Ion Batteries from. *Journal of The Electrochemical Society*, 9.

European Commission. (2023, August 17). *Circular economy: New law on more sustainable, circular and safe batteries enters into forc*. Retrieved from environment.ec.europa.eu: https://environment.ec.europa.eu/news/new-law-more-sustainable-circular-and-safe-batteries-enters-force-2023-08-17_en

European Parliament, C. o. (2023, July 28). REGULATION (EU) 2023/1542 OF THE EUROPEAN PARLIAMENT AND OF THE COUNCIL concerning batteries and waste batteries, amending Directive 2008/98/EC and Regulation (EU) 2019/1020 and repealing Directive 2006/66/EC. *Official Journal of the European Union*. Retrieved from eur-lex.europa.eu: https://celltech-group.com/press/eu-batteries-regulation/

EV BOX. (2023, 04 05). *Electric car battery weight explained*. Retrieved from blog.evbox.com: https://blog.evbox.com/ev-battery-weight

G. Towler, R. S. (2013). *Chemical Engineering Data.*

Group, C. (2024). *New EU Batteries Regulation – How should companies using batteries prepare in 2024?* Retrieved from Celltech-group.com: https://celltech-group.com/press/eu-batteries-regulation/

Li-Cycle Corp. Toronto, O. M. (2018). *Patent No. EP 3 535 803 B1.*

M. B. Rossi, S. v. (2024). *D6.2.1 Literature assesment of lithium ion battery recycling steps.* TNO. Utrecht: Green Transport Delta - Electrification.

M.S. Peters, K. T. (2003). *Plant Design and Costing for Chemical Engineers.*

Maxwell, C. (2020, May 28). *Cost Indices*. Retrieved from toweringskills.com: https://toweringskills.com/financial-analysis/cost-indices/

Northvolt AB, A. M. (2020). *Patent No. WO 2020/212363 A1.*

The European Green Deal. (2024, October). Retrieved from commission.europa.eu: https://commission.europa.eu/strategy-and-policy/priorities-2019-2024/story-von-der-leyen-commission/european-green-deal_en

Appendix

A. List of assumptions

Battery composition

Battery composition is based on two sources – the Northvolt patent (Patent No. WO 2020/212363 A1, 2020) and Diekmann et al. (2017) study (Diekmann, 2017). While Diekmann et al. (2017) gives a complete overview of a battery composition with the plastic casing and is easy to distinguish the composition of the cathode and anode materials and black mass in general (as presented in Table 1), in the Northvolt patent this was not the case. Therefore, it is assumed that the total weight percentage which sums up to 68.58% represents cathode active material, while the rest (31.42%) is assumed to be the battery casing.

Reaction stoichiometry

To close the material balance the reactions presented in Table 11 and their stoichiometry are considered:

Table 11: *Reaction stoichiometry as a basis for a material balance* (Patent No. EP 3 535 803 B1, 2018)

Leaching step	$6LiNi_{1/3}Mn_{1/3}Co_{1/3}O_{2(s)} + 9H_2SO_{4(aq)} + H_2O_{2(aq)} \rightarrow 2MnSO_{4(aq)} + 2NiSO_{4(aq)} + 2CoSO_{4(aq)} + 3Li_2SO_{4(aq)} + 2O_{2(g)} + 10H_2O_{(l)}$	(1)
	$Cu_{(s)} + H_2SO_{4(aq)} + H_2O_{2(aq)} \rightarrow CuSO_{4(aq)} + 2H_2O_{(l)}$	(2)
	$2Al_{(s)} + 3H_2SO_{4(aq)} + 3H_2O_{2(aq)} \rightarrow Al_2(SO_4)_{3(aq)} + 6H_2O_{(l)}$	(3)
	$2Fe_{(s)} + 3H_2SO_{4(aq)} + 3H_2O_{2(aq)} \rightarrow Fe_2(SO_4)_{3(aq)} + 6H_2O_{(l)}$	(4)
NMC cake precipitation	$CoSO_{4(aq)} + 2NaOH_{(aq)} \rightarrow Co(OH)_{2(aq)} + Na_2SO_{4(aq)}$	(5)
	$NiSO_{4(aq)} + 2NaOH_{(aq)} \rightarrow Ni(OH)_{2(aq)} + Na_2SO_{4(aq)}$	(6)
	$MnSO_{4(aq)} + 2NaOH_{(aq)} \rightarrow Mn(OH)_{2(aq)} + Na_2SO_{4(aq)}$	(7)
	$Li_2SO_{4(aq)} + 2NaOH_{(aq)} \rightarrow 2LiOH_{(aq)} + Na_2SO_{4(aq)}$	(8)
Al/Fe removal	$Al_2(SO_4)_3 + 2H_3PO_{4(aq)} \rightarrow 2AlPO_{4(s)} + 3H_2SO_{4(aq)}$	(9)
	$Fe_2(SO_4)_3 + 2H_3PO_{4(aq)} \rightarrow 2FePO_{4(s)} + 3H_2SO_{4(aq)}$	(10)
Neutralization and Li precipitation	$Na_2CO_{3(s)} + H_2SO_{4(aq)} \rightarrow Na_2SO_{4(aq)} + H_2O_{(l)} + CO_{2(g)}$	(11)
	$Na_2CO_{3(s)} + Li_2SO_{4(aq)} \rightarrow Li_2CO_{3(s)} + Na_2SO_{4(aq)}$	(12)
	$Na_2SO_{4(s)} + 10H_2O_{(l)} \rightarrow Na_2SO_4 \cdot 10H_2O_{(s)}$	(13)

System input

Solvent use

Sulfuric acid concentration is chosen to be the minimally sufficient.

In the leaching step, H_2SO_4 is used as a 2M H_2SO_4. The amount is following stoichiometry + 10% excess, as described in the Li-cycle patent (Patent No. EP 3 535 803 B1, 2018). In the same step, hydrogen peroxide is used as a 30 g/l aqueous solution in a stoichiometric amount.

In the solvent extraction step, there are several solvents used, always as a mixture of 70 wt.% kerosene and the rest are a solvent selective to a specific metal to be extracted from the pregnant leach solution (LIX860N, Cyanex 301, Cyanex 272, D2HPA). All these organic solvents are used as aqueous to organic ratio A:O=1:1.

In the back extraction, for stripping processes, H_2SO_4 is also used as aqueous solution of 2M H_2SO_4, and in ratio A:O=1:1. The sensitivity analysis is checking the impact of the A:O ratio in solvent extraction from 1:1 to 5:1, while in the back extraction it is checking from A:O is 1:1 to 1:5. This is all because of the most energy intense step in the solvent extraction route – concentrating highly diluted aqueous streams containing critical materials.

Solvent recovery

Solvent recovery is outside of battery limits. For the cost calculation, solvent loss is dependent on assumed equipment unit efficiencies except for LIX860N, where it is assumed that the loss is 0.1% of the one in the inlet stream. All losses are assumed to be replaced with virgin material (solvent). The solvent required for replenishing solvents that are required together with their costs are presented in Table 12 and Table 13 for mixed precipitation route and solvent extraction route respectively.

Table 12: *Overview of solvent amounts and costs needed for mixed precipitation route*

	Amount	Cost
H_2SO_4	46 kta	0.23 €/kg
H_2O_2	27 kta	0.31 €/kg
NaOH	4.22 kta	0.27 €/kg
H_3PO_4	1.02 kta	1.55 €/kg
Na_2CO_3	3.82 kta	0.27 €/kg
LIB860N*	0.11 kta	45.5 €/kg
Kerosene	0.27 kta	2.93 €/kg

**Cost for LIB860N isn't obtained, but assumed based on similar chemicals*

Table 13: *Overview of solvent amounts and costs needed for solvent extraction route*

	Amount	Cost
H_2SO_4	54.11 kta	0.23 €/kg
H_2O_2	26.44 kta	0.31 €/kg
H_3PO_4	1.02 kta	1.55 €/kg
Na_2CO_3	12.70 kta	0.27 €/kg
LIB860N	0.002 kta	45.5 €/kg
D2HPA	0.14 kta	1.1 €/kg
Cyanex 301	0.14 kta	1.8 €/kg
Cyanex 272	0.16 kta	1.8 €/kg
Kerosene	1.02 kta	2.93 €/kg

Process steps

Leaching

- The process takes place in a pressurized vessel
- Process efficiency 99.5%
- Solvents: 2M H_2SO_4 in stoichiometric amount + 10% excess; H_2O_2 in stoichiometric amount

Solvent extraction

Solvent extraction step consists of a mixer settler unit, a pump, a centrifuge where organic and aqueous phase are separated and a stripping unit where acid is again added for back extraction step.

Centrifuge efficiency is assumed to be 99.5%.

Stripping efficiency is assumed to be 99.8%.

In solvent extraction set of steps, all solvents used are a 30 wt.% solutions in 70 wt.% kerosene (D2HPA, Cyanex 301 and Cyanex 272 respectively).

Aqueous:Organic=1:1 (solvent extraction)

Aqueous:Organic=1:1 (back extraction)

Crystallization

Pressure vessels used.

Cost assumption: PVC vessels of approximately same cost as stainless steel.

Assumption: To reach the saturated solution for the crystallizer, streams have to be concentrated beforehand. Concentration by evaporation leads to forming of hydrated salts which then crystallize in the crystallizer. It is assumed that hot saturated solution enters the crystallizer at 80°C, after which the saturated mixture is cooed down to 30°C to crystalize metals in form of salts (Patent No. WO 2020/212363 A1, 2020). The products of crystallization are wet crystals and mother liquor. Wet crystals are sent to a filter where they are washed from mother liquor. After the mother liquor is removed, wet crystals (70 wt.%) containing water (30 wt.%) are sent to drier where they are dried and also dehydrated to their anhydrous forms. For the process description, data on solubility is used. Where the information was missing, interpolation and extrapolation are used and corresponding assumptions on solubility are made.

Filtration

It is assumed that after crystallization, mother liquor is removed by washing and filtration, while wet crystals contain 10 wt.% of mother liquor.

Washing

It is assumed the crystals are washed during the filtration step. It is assumed that 1 kg of water is needed per 1 kg of crystals. It is assumed that now wet crystals contain 30 wt.% water which is removed in the drying step.

Costs

Once the material and energy balance are closed, the next step in the economics is sizing of the equipment as described in the process scheme. Equipment size is the starting point for the evaluation of the total equipment cost, which is then corrected for the CEPCI index 2023 (797.9). All values estimated in economic analysis are based on data from “Chemical Engineering Design” (G. Towler, 2013) and “Plant Design and Costing for Chemical Engineers” (M.S. Peters, 2003). For this type of process, defined as a solid-liquid process new process, a novelty factor of 10% is added on the equipment cost, resulting in Total Equipment Cost (TEC).

Total delivered cost adds the delivery of purchased equipment factor (10%) of TEC to TEC. **Total delivered cost is the basis** for the estimation of total direct cost (TDC), Total indirect cost (TIC) and working capital (WC). The overview of total direct cost and total indirect cost together with factors used in calculation is presented in Table 19 and Table 20 respectively.

Costs of consumables and utilities for two routes are presented in Table 21 - Table 24.

Total fixed capital investment (FCI) summarizes total direct and total indirect costs, while total capital investment (TCI) represents the sum of working capital and total fixed capital investment (FCI).

Fixed capital investment (FCI) is further used as a basis for estimation of fixed and variable production costs alongside raw material costs, labour costs, utilities and consumables cost. Total production costs without depreciation (TPCWD) are estimated as a sum of fixed and variable production costs with plant overhead and administration costs. These are represented in Table 25, Table 26 and Table 27.

B. Techno-economic analysis

1.1 Material and Energy Balances

The following assumptions are made to mimic a commercial scale plant:

Aqueous:Organic ratio = 1:1 in solvent extraction.

Aqueous:Organic ratio = 1:1 in back extraction.

Solvent losses are assumed to depend on individual process step efficiency and are assumed to be covered by adding virgin materials, whose quantities are shown in Table 12 and Table 13.

Solvents for solvent extraction are assumed to be used endlessly.

Energy balances were calculated based on mass balances. The plant energetic needs are presented in Table 5. The energy needs are split into three categories: heat, cooling and electricity.

Using heat exchangers, heating and cooling needs are assumed to be integrated up to 80%.

All heating needs are provided by saturated steam at 790 kPa.

1.2 Equipment Sizing

The major process equipment envisioned in the plant design and its characteristic size (used for cost estimation) is presented in Table 14 for mixed precipitation route and Table 15 for solvent extraction route. Based on the calculated sizing variables depending on the process unit, the cost of the equipment is estimated and presented in Section 1.4.1.

Table 14: *Equipment cost for the mixed precipitation route*

Equipment type	**Equipment**	**Sizing variable**	**Value**	**Unit**	**Number**	**Equipment cost, M€**
Storage Tanks	cone roof	capacity	35	m^3	1	0.051
Storage Tanks	cone roof	capacity	25	m^3	1	0.043
Storage Tanks	cone roof	capacity	10	m^3	11	0.310
Heat Exchangers	U-tube shell and tube	area	16	m^2	2	0.120
Agitators & mixers	Propeller	driver power	18	kW	1	0.083
Filters	Vacuum drum	area	10	m^2	1	0.230
Pressure vessels	Vertical, cs	shell mass	4000	kg	1	0.104
Agitators & mixers	Propeller	driver power	20	kW	1	0.088
Dryers	Direct contact Rotary	area	11	m^2	1	0.216
Pressure vessels	Vertical, cs	shell mass	4400	kg	1	0.110
Agitators & mixers	Propeller	driver power	22	kW	1	0.094
Dryers	Direct contact Rotary	area	11	m^2	1	0.216
Pressure vessels	Vertical, cs	shell mass	418	kg	1	0.035
Dryers	Direct contact Rotary	area	11	m^2	1	0.216
Conveyors	Belt, 0.5 m wide	length	10	m	2	0.197
Pressure vessels	Vertical, cs	shell mass	3500	kg	1	0.095
Distillation columns	Distillation columns	length	1.5	m	1	0.058
Pumps and drivers	Single stage centrifugal	flow	3	liters/s	4	0.071
Pumps and drivers	Single stage centrifugal	flow	6.2	liters/s	4	0.075
Storage Tanks	cone roof	capacity	72	m^3	1	0.077
Unlisted equipment	Electrowinning (Allanore, 2020)	cost	3348800	€	1	5.181
Total equipment cost (CEPCI 2023)						**M€ 7.668**

Table 15: *Equipment cost for the solvent extraction route*

Equipment type	**Equipment**	**Sizing variable**	**Value**	**Unit**	**Number**	**Equipment cost, M€**
Storage Tanks	cone roof	capacity	10	m^3	10	0.282
Storage Tanks	cone roof	capacity	30	m^3	5	0.236
Storage Tanks	cone roof	capacity	40	m^3	1	0.055
Storage Tanks	cone roof	capacity	21	m^3	1	0.039
Heat Exchangers	U-tube shell and tube	area	22	m^2	2	0.123
Agitators & mixers	Propeller	driver power	21	kW	1	0.091
Heat Exchangers	U-tube shell and tube	area	16	m^2	1	0.060
Filters	Vacuum drum	area	10	m^2	1	0.230
Storage Tanks	cone roof	capacity	46	m^3	16	0.951
Agitators & mixers	Propeller	driver power	46	kW	8	1.304
Storage Tanks	cone roof	capacity	20	m^3	1	0.038
Agitators & mixers	Propeller	driver power	20	kW	1	0.088
Dryers	Direct contact Rotary	area	13	m^2	1	0.246
Pumps and drivers	Single stage centrifugal	flow	5	liters/s	2	0.037
Storage Tanks	cone roof	capacity	51	m^3	16	1.008
Agitators & mixers	Propeller	driver power	51	kW	8	1.422
Pumps and drivers	Single stage centrifugal	flow	5.6	liters/s	4	0.074
Evaporators	Vertical tube	area	60	m^2	6	4.190
Crystallizers	Crystallizers	capacity	1	t/h	3	1.340
Agitators & mixers	Propeller	driver power	5	kW	3	0.043
Filters	Vacuum drum	area	10	m^2	3	0.173
Dryers	Direct contact Rotary	area	11	m^2	2	0.432
Conveyors	Belt, 0.5 m wide	length	10	m	1	0.098
Pumps and drivers	Single stage centrifugal	flow	1	liters/s	6	0.101
Pumps and drivers	Single stage centrifugal	flow	2.5	liters/s	9	0.157
Pumps and drivers	Single stage centrifugal	flow	3.5	liters/s	5	0.089
Unlisted equipment	Electrowinning (Allanore, 2020)	cost	3348800	€	1	5.181
Total equipment cost (CEPCI 2023)						**M€ 18.087**

Minor equipment (like some heat exchangers, pumps, vacuum pumps, screw conveyers, etc.) are not listed as they fall within the error margin of the estimate resulting from this analysis.

1.3 Economic Analysis

Based on the mass and energy balance conducted in the previous section, the economic analysis has been performed. This is split into equipment costing, material costing, utility costing and an overview. A number of assumptions had to be made regarding the material, energy and operation costs. These are highlighted in grey blocks.

1.3.1 Capital Investment

Equipment Costs

Based on data and correlations from literature2,3, and by using CEPCI indexes (Maxwell, 2020) from Table 16, Table 17 and Table 18 the total equipment cost has been generated and adjusted for the calculation year (2023) and currency (€). Additionally, the novelty factor of 10% (Table 17 and Table 18) is included in the equipment costs estimation. Assuming that the site is not entirely prepared for delivery of such equipment, additional delivery costs of 10% are included. Based on this total equipment delivery cost subsequent direct and indirect capital expenditures are calculated.

Table 16: *CEPCI index* (Maxwell, 2020)

Year	CEPCI
2010	550.9
2011	585.7
2012	584.6
2013	567.3
2014	576.1
2015	556.8
2016	541.7
2017	567.5
2018	603.5
2019	619.2
2021	686.0
2022	699.0
2023	797.9
2024	800.7

Table 17: *Total equipment cost (TEC) for the mixed precipitation route*

TEC reference year of 2010 (CEPCI: 550.9)[4]	M€	1.718
TEC calculation year of 2023 (CEPCI: 797.9)	M€	7.668
TEC including novelty factor (10%)	M€	8.435

Table 18: *Total equipment cost (TEC) for the solvent extraction route*

TEC reference year of 2010 (CEPCI: 550.9)[3]	M€	8.911
TEC calculation year of 2023 (CEPCI: 797.9)	M€	18.087
TEC including novelty factor (10%)	M€	19.896

---

[2] 'Chemical Engineering Design' by G. Towler, R.K. Sinnott, 2nd Ed, 2013

[3] 'Plant Design and Costing for Chemical Engineers', M.S. Peters, K.D. Timmerhaus, R.E. West, 5th Ed, 2003

[4] This excludes electrowinning cell as it comes *from Stinn, Caspar and Antoine Allanore. "Estimating the Capital Costs of Electrowinning Processes." The Electrochemical Society Interface 29, 2 (June 2020): 44. © 2020 The Electrochemical Society*

Table 19: *Direct costs.*

Cost	**Mixed precipitation**	**Solvent extraction**
	Amount M€	
Total delivery cost	9.279	21.885
Purchased equipment installation (39%)	3.619	8.535
Instrumentation & Controls (installed) (26%)	2.412	5.690
Piping (installed) (31%)	2.876	6.784
Electrical systems (installed) (10%)	0.928	2.189
Buildings (including services) (29%)	2.691	6.347
Yard improvements (12%)	1.113	2.626
Service facilities (installed) (55%)	5.103	12.037
Total direct cost	**18.743**	**66.094**

Table 20: *Indirect costs.*

Cost	**Mixed precipitation**	**Solvent extraction**
	Amount M€	
Engineering and supervision (32%)	2.969	7.003
Construction expenses (34%)	3.155	7.441
Legal expenses (4%)	0.371	0.875
Contractor's fee (19%)	1.763	4.158
Contingency (37%)	3.433	8.098
Total indirect cost	**11.691**	27.575

Total Capital Investment (TCI) is depreciated over 5 years using the straight-line method.

Production Costs

Production costs are composed of material, labor, energy and other overhead costs.

Material Costs

Table 21: *Consumable costs for mixed precipitation route*

Material Name	**Unit Price** **€/t**	**Annual Amount** **t/y**	**Annual Cost** **M€**
H_2SO_4	227.5	46000	10.465
H_2O_2	309.4	27000	8.354
NaOH	273.0	4220	1.152
H_3PO_4	1547	1020	1.578
Na_2CO_3	273.0	3820	1.043
LIB860N	45500	11.46	0.521
Kerosene	2939.3	26.74	0.079
		Total Consumable Cost	**23.192**

Table 22: *Consumable costs for solvent extraction route*

Material Name	**Unit Price** **€/t**	**Annual Amount** **t/y**	**Annual Cost** **M€**
H_2SO_4	227.5	54113	12.311
H_2O_2	309.4	26444	8.182
H_3PO_4	1547	1020	1.578
Na_2CO_3	273.0	12701	3.467
LIB860N		2	0.093
Cyanex 301	2000	142.2	0.259
Cyanex 272	2000	160	0.291
D2HPA	1200	1021	0.147
Kerosene	3232	135	3.003
		Total Consumable Cost	**29.331**

Labor Costs

The following assumptions were made regarding labor costs:

Number of workers: 2

Number shifts: 3

Number of days per year: 330

Hourly rate: 40 €/hr

For the given assumptions, total annual labor cost are **0.50 M€** and this is valid for both processes**.**

Utility Costs

Table 23: *Total utility costs (TUC) for mixed precipitation route.*

Utility Name	Unit Cost	Cost Unit	Annual Requirement	Req. Unit	Utility Cost
					M€/year
Electricity	0.045	**€/kWh**	934 740	kWh/y	0.042
Water	0.0045	**€/m^3**	33 789	m^3/y	0.0002
Sat. steam (790 kPa)	23.20	**€/t**	22 345	t/y	0.518
Total Utilities Cost					**0.561**

Table 24: *Total utility costs (TUC) for solvent extraction route.*

Utility Name	Unit Cost	Cost Unit	Annual Requirement	Req. Unit	Utility Cost
					M€/year
Electricity	0.045	**€/kWh**	6 900 500	kWh/y	0.314
Water	0.0045	**€/m^3**	76 500	m^3/y	0.0003
Sat. steam (790 kPa)	23.20	**€/t**	64 000	t/y	1.426
Total Utilities Cost					1.741

Operation Costs

Table 25: *Variable production costs.*

Cost Item	Mixed precipitation	Solvent extraction
	Cost M€/year	
Raw material	4.368	4.368
Consumables	23.192	29.331
Total Labour Costs (TLC)	0.490	0.490
Operating Supervision (OS) (15% TLC)	0.074	0.074
Utilities	0.561	1.741
Maintenance & Repairs (M&R) (6% FCI)	2.168	5.620
Operating Supplies (15% M&R)	0.325	0.843
Laboratory Charges (15% TLC)	0.074	0.074
Variable Cost of Production	**31.252**	**42.541**

Table 26: *Fixed production costs.*

Cost Item	Mixed precipitation	Solvent extraction
	Cost M€/year	
Taxes (property) (2% FCI)	0.794	1.873
Insurance (1% FCI)	0.397	0.937
Fixed Cost of Production	**1.191**	**2.810**

Table 27: *General production costs.*

Cost Item	Mixed precipitation	Solvent extraction
	Cost M€/year	
Administration (20% TLC+SVC+M&R)	0.546	1.237
Plant Overhead (60% TLC+SVC+M&R)	1.639	3.710
Distribution & Selling (5% TPC)	1.924	2.794
Research & Development (4% TPC)	1.539	2.235
Royalties (if not lump sum) (1% TPC)	0.385	0.559
General Production Costs	**6.033**	**10.535**