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Conceptual Process Design

Process Systems Engineering
DelftChemTech - Faculty of Applied Sciences
Delft University of Technology

Subject

DESIGN OF AN INTEGRATED FERMENTATION CRYSTALLIZATION PROCESS FOR THE PRODUCTION OF L-PHENYLALANINE

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PREFACE

This report describes the design of integrated fermentation-crystallization process for production of L-Phenylalanine. It has been the design team's purpose to identify and overcome the design challenges and to come to a good design. Considering these goals, an ISPR (In-Situ Product Removal) process was designed.

While overcoming the challenges, the team was expected handling of uncertainties and coming up with solutions creative and technically feasible. In the point of decision, team asked help of the coaches and received the required feedback. The present final design based on decisions, which were made carefully and approved by every member of the team. We hope that the work done will meet the expectations.

During the preparation of this study, we have consulted our coaches many times for the problems we have faced. Here for the first time we find the opportunity to thank our coaches for their support and their understanding. We would like to thank our supervisor, Dr.Ir.Adrie Straathof and our principal PDEng.Maria Cuellar for all the information and support they gave us regarding the design. Furthermore we would like to thank our creativity coach, Prof.Henk Nugteren, for sharing his time with us for our project and for our problems.

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October 19th, 2005
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SUMMARY

This work is a part of the course Conceptual Process Design. The purpose of this work is to design a plant for the production of the amino acid L-Phenylalanine. The desired goal is to achieve a production of 1000 t/a of the product with a purity of 98.5%. The required presentation of the product is the anhydrous form of its crystals due to the undesired properties of the monohydrate crystals.

Several obstacles had to be taken in account for the development of the process such as product inhibition form the target compound on the microorganism, product crystallization on the fermentor due to the low solubility of the product and the use of as less as possible units for the downstream process.

The approach for this design is the ISPR with integrated fermentation crystallization processes. This approach is used to lower the concentration of the target product and hence to overcome the problems mentioned before. Characteristics of the product had to be taken into account for the other parts of the process like the temperature for the different parts of the process in order to maintain the product in good conditions.

By taking account the different factors like productivity and economics some improvements are considered and presented with this report. The designed process fulfills the constraints set by the client.

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PREFACE

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1. INTRODUCTION

The (bio) chemical industry nowadays is evolving in to the production of different compounds traditionally made by chemical means to the use of fermentation processes with recombinant microorganism. These processes are characterized by the low concentration of the product in the fermentation broth making the separation difficult and costly. The downstream process is crucial for the design of the production facilities affecting the production, purity and economics of the process. Another important aspect to consider is the conditions that have to be met for the desired production process and the effect that different compounds might have on on the units and their performance like the interaction with the microorganism, side products formation and the separation processes required to obtain the specified product. One approach done in the TU Delft is the implementation of an ISPR (In situ product removal) in fermentation processes to overcome possible problems like product inhibition or toxicity, enhance the production of the desired compounds and improve the performance of the different steps of the overall (Buque-Taboada *et al.* 2004).

The objective of this work is to do a conceptual process design of and L-Phenylalanine production plant with a capacity of 1000 ton/yr of the product with a purity of 98.5% using an ISPR process with a recombinant strain of *E.coli*, This compound is of industrial and pharmaceutical importance and has special properties SEE APENDIX 4.3 that have to be taken in consideration to obtain the production level and quality for the market

Traditionally L-Phenylalanine is made by different processes like the resolution of N-acetyl-D,L-Phenylalanine or the Stereo-and enantioselective addition of ammonia to trans-cinnamic acid. These processes are chemical and the product obtained is D,L-Phenylalanine that has to be separated increasing the steps for the downstream processing, the wastes and raw materials used, making the development of a fermentation process with a recombinant strain of a glucose consuming, L-Phenylalanine producing microorganism an interesting subject for research. For the reasons stated before and from an initial perspective this process might be economically more advantageous, thus, research has focused on the production of L-Phe with isolated enzymes, immobilized cells, resting cells using different microorganisms like *Corynebacterium glutamicum*, *Brevibacterium flavum* and most importantly *E. coli* (Takors, 2004). Further more the addition of several recycle streams to the initial scheme proposed by the principal will decrease the materials used as well as the wastes.

The aromatic amino acid L-Phenylalanine (L-Phe) is of pharmaceutical importance for parenteral nutrition and represents an important building block for synthesis of aspartame, which is a low-calorie sweetener and widely used in various foods and dry beverage mixes owing to its approximately 200-fold stronger sweetness compared to that of sucrose (Takors, 2004). The worldwide L-Phenylalanine production is estimated to be 14,000 tons in 2002 (Budzinski, 2001). The market price of L-Phe is 20 to 24 US$/kg for chemical or 30 to 40 US$/kg for pharmaceutical (Bongaerts *et al.*, 2001).

Table 1.1 L-Phenylalanine producers.[1]

Company	Location
Absolute Standards, Inc	Andover, MA, 01810-5413, USA
ANDAChem, Inc	Taiyuan, People's republic of China
Acesys Pharmatech, Ltd.	Oradell, NJ, 07049, USA
Magical Scientific	Oklahoma City, OK, 73156-1326, USA
DSL Chemicals (Shanghai) Co., Ltd.	Shanghai, People's Republic of China

The production process and lab scale ISPR has been studied by several researches like Maass *et al*, 2002 using reactive extraction; Kusunose *et al.*, 2004 using polymeric beads and by R. Takors 2004 (Chapter 2.1). Currently in TU Delft, PDEng. Maria Cuellar is doing research in the production of L-Phenylalanine with ISPR by means of an integrated fermentation crystallization process in order to overcome some of the problems mentioned above like product inhibition, crystallization of the product in the fermentor and reduction in downstream processing.

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[1] SciFinder Scholar

2. PROCESS OPTIONS AND SELECTION

This chapter contains the different ISPR process options for the production of L-Phenylalanine with the block schemes for each one.

2.1. Process Concept Chosen

Criteria and impact factors

In the process case assigned, an ISPR phenylalanine production process is described, using the fermentation conditions described by Gerigk *et al*, 2002 and the research done by the principal of the project. Different ISPR processes found in literature were used to compare the process suggested taking into account the objectives formulated and that an integrated fermentation crystallization process has not been studied (see table 2.1).
Further more some changes to the given process were made in order to improve the use of the raw materials and minimize the wastes that will be explained further.

The comparison between the different ISPR processes and the selected process will be according to the following criterions:

Table 2.1. Criteria and impact factors

Criterion	Impact factor
Purity (Yes/No)	Requirement
Auxiliary Materials	4
Number of Units	4
Yield	5
Control	3
Environmental Safety	4

- Purity: <Yes/No> The process chosen has to fulfill the purity constraint of product, which is 98.5%. Purity is of major importance because the product will aim for an industrial and not pharmaceutical application affecting the price of the product in a high percentage.

- Auxiliary Materials: The requirement of higher amount of auxiliary materials will increase the production cost as well as the waste. The use of auxiliary materials can also affect the number of equipments used in the process constituting one of the most important objectives of the research.
- Number of Units: The number of units will increase the investment cost of the designed process. This aspect is related greatly to the purity of the product, quality and cost of the product and process.
- Yield: Higher yield is an important criterion for the selection of the process.
- Control: Any control error may lead loss of money or even loss of lives. Therefore control of process is desired to be secure and easy.
- Environmental Safety and Waste Treatment: The process chosen has to obey to the standard environmental regulations to protect the environment and indirectly the human health.

Yield is thought to be most important criterion because high yield will pay the possible additional cost of high technology and equipment costs although it may be seen unfavorable due to the high investment requirement. Secondly, control is an important criterion; the performance and quality of the integrated fermentation crystallization process depends on it. Environmental safety is and important aspect to consider because of the use of recombinant strains and the legislation regarding this kind of processes, especially the wastes regarding the recombinant strain have to be handled carefully and disposed in a safely manner. The process must have a good use of the materials in order to minimize the wastes and treatment of the same. On the other hand, environmental policy of DSM is well known and respected in industrial and public level. When DSM is concerned, environmental safety is one of the most important aspects taken into account in process evaluation.

The first obligation of each process is the achievement of high purity demand of the client (98.5%). This purity demand of the client also gives information about the target

use of the product. L-Phenylalanine with very high purities like higher than 99.5% is used in the pharmaceutical industry and in medicine. Lower purities like 98.5%, however, are good enough for the commercial use as artificial sweetener (as raw material for aspartame or as additive in food industry). Therefore the target of the client is industrial chemical use.
Because of this reason the processes yielding higher purities do not gain any priority in the selection. Hence the condition is a yes-or-no question and all the proposed processes fulfill this requirement.

Auxiliary materials consist of the subsidiary materials like solvents, membranes, carriers and additives that are needed in some of the evaluated processes. Usually the use of auxiliary materials increases the cost of the operating process and affects the downstream processing as well as an increase in the wastes.

Among the alternative processes the first process is the most advantageous one due to less use of auxiliary materials and downstream processing units. Second process utilizes solvents and runs a solvent-based crystallization operation. On the other hand the extraction unit uses membranes where they should be replaced each time after certain use. Third alternative uses beads, which is expensive despite being cheaper than the use of ion exchange column (scheme 4), as claimed in the article written by Kusunose, 2004.

Operating time

The production requirement for the process is 1000-ton annual production of L-Phenylalanine crystals. Because of the lack of information on a industrial process similar to the proposed and the useful life of the biomass, the batch time is set to be 363 hours[2] the reason is that the production process in biotechnology are not economically feasible for shorter production time, especially for the desired specifications of the product. In the studies of Takors and Gerigk, the fermentation is stopped after 50 hours due to the increasing acetate formation in the process. However in the same studies the information about the acetate formation is not given. Therefore the acetate formation had to be

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[2] By recommendation of the Technical Supervisor

assumed with a reasonable value related to the glucose fed to the system. Therefore it is assumed that 0.5% of the glucose fed to the system is consumed for the acetate production, this is explained by considering very good mixing and high levels of dissolved oxygen in the fermentor and the biomass recycle in order to prevent the anaerobic acetate production. However, even with the control of the glucose present in the system and oxygen in excess, acetate production took place, meaning that the acetate production is not because of the overflow metabolism of glucose or oxygen depletion, but it certainly reduces it. In the calculations based on these assumptions acetate concentration was not a problem and the production phase can be done successfully for the 2 weeks of period.

2.2.In Situ Product Removal (ISPR)

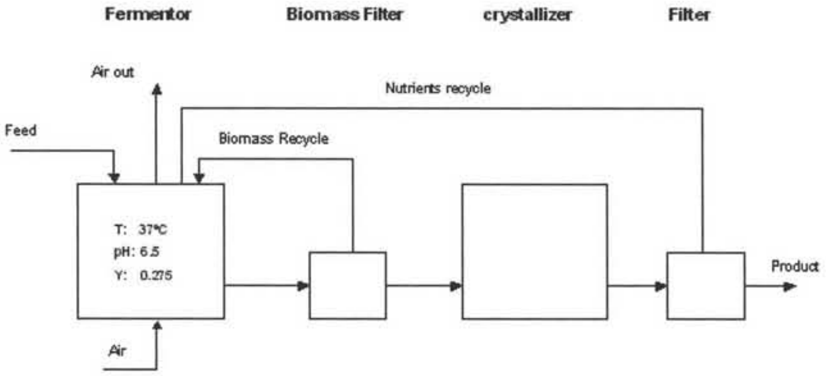
Several reported processes have been evaluated for the ISPR of L-Phenylalanine. The differences in ISPR methods and the effect it has on the rest of the process were evaluated using the criterions mentioned in Table 2.1. In all the processes the product removal is made outside the fermentor to avoid complex equipments and reduction in their performance.

Before describing each process separately, it might be very useful to mention about the main common properties and differences of them. First of all, each process uses ISPR approach for the production of L-Phenylalanine. The product is taken out from the fermentor during the production phase, whereas biomass recycled to the fermentor with the use of a separation unit. In the latter three processes there is found one additional unit after the biomass separation process to transfer the L-Phenylalanine from the fermentation medium to crystallization medium. This operation occurs in three different ways: i) Through a reactive extraction process where carriers are used (Ruffer *et al*, 2004), ii) With the use of beads inside the fermentor medium and afterwards through washing with water in a column (Kusunose and Wang,2004) , iii) Through Ion-Exchange Column where the isoelectric property of the L-Phenylalanine is used for the separation.

The fermentation broth returns to the fermentor after the separation process, whereas liquid broth carrying L-Phenylalanine flows to the down-stream crystallization unit. In this point it is very important to mention that the crystallization is not a part of ISPR process, but a part of the down-stream processing.

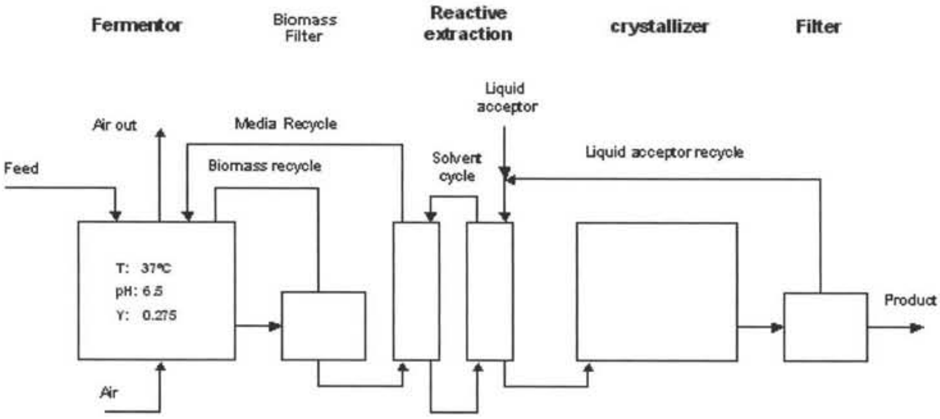
The first process evaluated is the proposed by the principal consisting of the integrated fermentation-crystallization production of the L-Phe (Figure 2.1). This process has the advantage that the goal of reducing the concentration of the target compound, maintaining a relative high production rate and reducing the negatives effect caused by the product is met while simplifying the downstream processing. This process was selected as the base case after evaluating the criterions mentioned above. The other alternatives use different ISPR processes where the final result is solution containing the L-Phe so a crystallization step is generally needed and its downstream processing to obtain the final product.

Figure 2.1 In-Situ Crystallization-Fermentation (Scheme 1)



- The second alternative process is In-Situ Reactive Extraction-Fermentation Process, is seen in (Figure 2.2). The reactive extraction process described by Maass *et al.* 2002 and Takors, 2004, uses extraction unit consisting of two hollow fiber modules, one is used for extraction and the other for back extraction. Extraction is realized by using kerosene as the organic solvent with a cation selective carrier DEHPA and sulfuric acid as the proton counter-ion donor in the liquid acceptor phase, achieving an increase on the yield of 35% compared to the non-ISPR process. First the broth is filtered to remove the biomass and pumped to the extraction unit. The aqueous solution containing the L-Phenylalanine is brought in contact with the organic phase consisting of a kerosene solution and the cation selective carrier DEHPA (10% v/v). The ion exchange takes place in the aqueous/organic interface. The organic phase with the carrier/L-Phenylalanine complex is transported to the back extraction unit where the L-Phe$^+$ Is replaced by H$^+$ from the H_2SO_4 solution by contact with the organic phase. The aqueous phases are pumped through the hydrophobic hollow fibers and the organic phase circulates in the opposite direction between the modules using the outer fiber space. The stripping phase containing the extracted L-Phe is then taken for the downstream processing to remove the product.

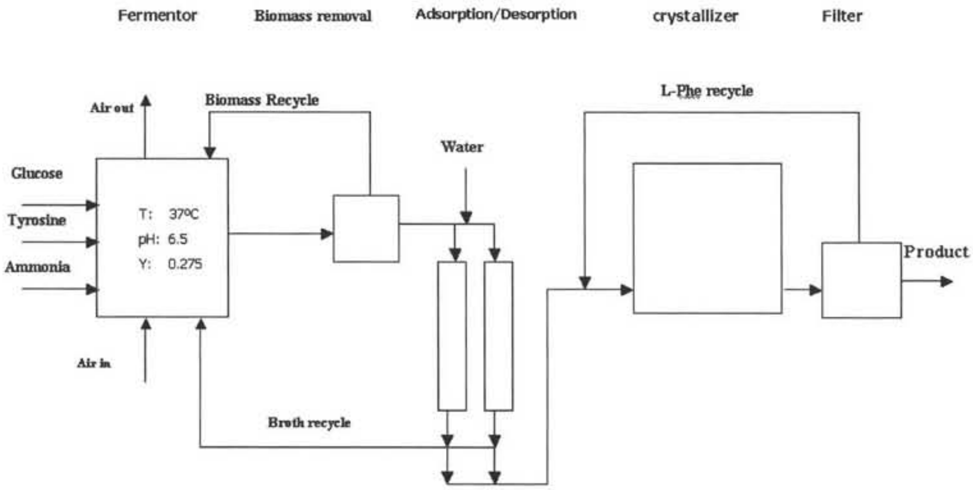
Figure 2.2 In-Situ Reactive Extraction- Fermentation Process (Scheme 2)



This process has the disadvantage that the extraction and back extraction units need auxiliary materials and specialized equipment. This might increase the materials used, the waste, cost for operating the process and the downstream processing of the product stream. Also the product obtained with this method has a very high purity, more than 99% (Maass *et al.* 2002). The purity required by the client is lower than the obtained by this method, so the product is not useful.

- In-Situ Adsorption-Fermentation Process was the third option considered with the use of beads as seen in Figure 2.3. In order to increase the production and avoid feedback inhibition and crystallization in the fermentation process, an inert polymeric adsorbent is used to lower the concentration of the product in the fermentor and avoid the crystallization of the target compound inside the vessel. When the production phase starts the product from the fermentor is separated from the biomass, and recycled to the fermentor The polymeric beads are located in two parallel columns in order to have a continuous product removal in one of the columns while the other one is being stripped from the L-Phe attached to the beads. L-Phenylalanine is desorbed using simple water since the adsorption interaction between the polymeric beads and the L-Phe is weak (Kusunose, 2004). The liquid resulting from the adsorption step in the column is then recycled back to the fermentor. The water used for desorption of the target compound is taken to the downstream processing units to obtain the product in the desired form. After crystallization, the water with diluted L-Phe is recycled back to the crystallizer to recover as much L-Phe as possible.

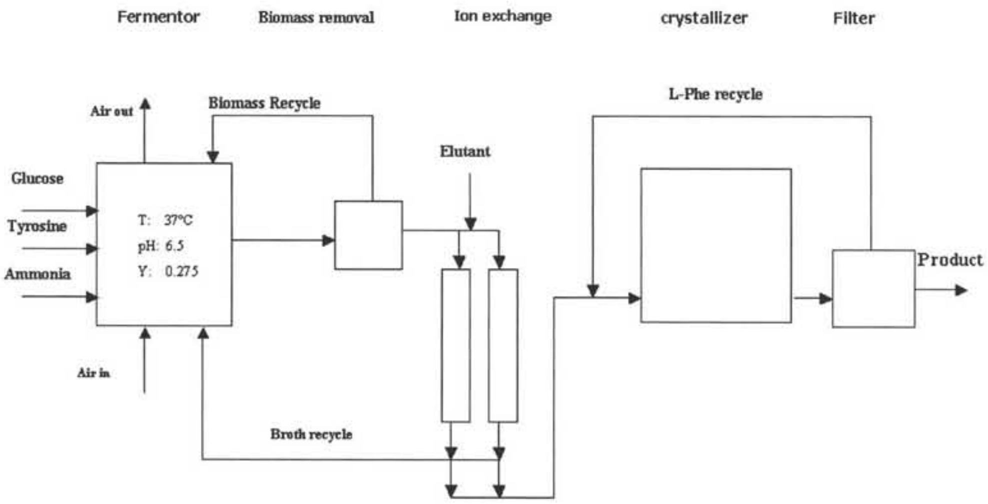
Figure 2.3 In-Situ Adsorption-Fermentation Process with the Use of Beads (Scheme 3)



In studies for the production of L-Phenylalanine made by Kusunose and Wang, 2004; using polymeric beads inside the fermentor and a modified strain of *Brevibacterium lactofermentum* showed that 40% of the L-Phenylalanine produced was extracted during fermentation proving to be interesting method of ISPR improving the overall fermentation performance.

- The fourth option is an unpublished design project of Chemical Engineering Design Group of West Virginia University, Figure 2.4. This design involves the batch production of four types of amino acids: L-Aspartic acid, L-Phenylalanine, L-Lysine HCl, and L-Leucine. The common property of these amino acids is that they primarily used as dietary supplements. The referred process is a batch process, however it can easily be modified to in-situ operation. The process then will be similar to scheme 3 but with the difference that the pH of the fermentation broth after biomass removal must be altered in order to obtain the L-Phe in the ionic form. The stream is taken to an ion exchange tower where the amino acid is removed using an ion exchange resin (Dowex Marathon C). To work continuously, a set of two ion exchange columns are used so while one is in use, the other one is being stripped with an elutant. (Reference: website of west Virginia University, www.che.cemr.wvu.edu/publications/projects/large_proj/batch-production_of_amino_acids.pdf)

Figure 2.4 In-Situ Ion Exchange-Fermentation Process (Scheme 4)



This process has the disadvantage that needs more materials in order to alter the pH, and extract the L-Phe from the ion exchange column; different to scheme 3 where the elutant was water so the waste treatment would be easier. Also the ion exchange slurry needs to pass trough at least a crystallization unit to obtain the desired product. The advantage of this process is that is flexible and it can be used to produce different kinds of amino acids with the same equipment.

In Table 2.2, the four different processes are evaluated according to the criterion mentioned in Table 2.1 and compared to each other. The results indicate that the first option is the most suitable for the development of the process.

Table 2.2 Process Evaluation Table

Criteria\Concept	**Impact factor**	**Scheme1**	**Scheme2**	**Scheme 3**	**Scheme 4**
Purity (Yes/No)	Requirement	Yes	Yes	Yes	Yes
Auxiliary Materials	4	+	--	0	--
Number of Units	4	+	-	-	-
Yield	5	n.a	++	+	+
Control	3	-	0	-	-
Environmental Safety	4	+	0	+	0
TOTAL		9	2	2	-10

"0" For neutral

"+" Advantage

"-" Disadvantage

"n.a" No information

2.3. Conclusion

Taking as reference the process evaluation in Table 2.2 the process to be developed will be the scheme given by the client consisting on a fermentation-Crystallization process. The final process will vary in comparison to the proposed due to several recycle streams added as seen in Appendix 3.3, to improve the performance, the use of the raw materials and the design of the case.

3. BASIS OF DESIGN (BOD)

In this chapter the selected option to be developed further will be presented along with the modifications that were made to reach the objectives desired by the clients and developed by the designing team

3.1. Description of the Design

An industrial process for an annual production of 1000 ton of anhydrate L-Phenylalanine crystals with a mean particle size of 100μm and a purity of 98.5% is developed. The process consists of a fermentation crystallization process with a modified strain of *E. coli* as the catalyst that has the aromatic amino acids biosynthetic pathway altered to produce more L-Phenylalanine. Several product characteristics have to be taken in account for the process such as, product inhibition at low concentrations, low solubility of the product that might result in crystallization in the fermentor and the polymorphism of the crystal form of the product. To deal with the inhibition and solubility (crystallization) problem inside the fermentor an In-Situ Product Removal fermentation crystallization process is developed taking as reference the studies made by Gerigk *et al*, 2002, Maass *et al*, 2002 and Takors *et al*, 2004. The product market price is 20 to 24 US$/kg (Bongaerts *et al.*, 2001) and the goal of the process is to obtain a profit with a product price of around 10 US$/kg.

3.2. Process Definition

For the decisions of the production process to follow, different processes were evaluated and one was selected as a base case to develop further (Chapter 2.1.1). A summary with the characteristics of the process, block diagram of the selected case, the complete case, important kinetic information and thermodynamic properties of the components will be stated.

3.2.1. Process Concept Chosen

In Chapter 2 the processes considered were analyzed. According to the criterion described there. After the qualitative evaluation, the best option was chosen, as seen in Table 2.2.

For the reactive extraction, polymeric beads and ion exchange columns, more materials must be used and as consequence the downstream processing is more complex, in some cases the purity is higher tan required, more units are needed and the wastes are higher because of the use of different solvents, carriers etc. The process chosen has the lowest possible downstream process, the materials needed are the same for the fermentation process in the other alternatives, but the downstream processing is made with the use of water which simplifies greatly the purifying of the product and the equipment needed

Further developments were made to the process chosen taking in consideration the characteristics of each unit operation in order to achieve the best possible configuration with the least amount of looses.

3.2.2. Block Schemes

Selected process

The process block scheme chosen for the basis of the design can be seen on Figure 3.1. The information in the diagram contain the mass flow of the streams, the operating temperature and pressure of the equipments as well as the preliminary assumption for the yield of each unit, the yield for the fermentation unit was found using the kinetics provided by the client and the information given in the studies made by Gerigk, *et al*, 2004; the overall yield for the initial process chosen is18% with an annual production of 1000 tons in approximately 8000 hours.

The preliminary process results and the time duration for each phase was assumed to be the same as the ones observed by Rüffer *et al*, 2004 in his experiments. This time is considered to be very short and is due to the decrease in production of L-Phenylalanine and acetate accumulation. The time duration for each phase is shown in table 3.1a. and Table 3.1.b for the process developed y Rüffer *et al*, 2004 and for the designed process respectively.

Table 3.1a Indication of operation time based on Rüffer *et al*, 2004.

Unit operation		Process time (h)
Fermentation[1]	Batch phase	7
	Extension of the growth phase	7
	Production phase + crystallization[2]	36
Rest time[3]		5
Batch time		55
Working time per year		8000
Batches per year		146

[1] From literature in the studies made by Gerigk. et. al, 2002 and Rüffer *et al*, 2004.

[2] Time of crystallization used during the L-Phe production phase. 15 hour is the estimated value; the accurate value will be worked out in the final report.

[3] Include drying, filling, cleaning, and etc. It is estimated value.

Table 3.1b Indication of operation time for the definitive case.

Unit operation		Process time (h)
Fermentation	Starting phase[1]	1
	Batch phase	9
	Fed Batch until product Removal	12.3
	Steady state production phase	336
Rest time[2]		52.4
Batch time[3]		408.7
Working time per year		8000
Batches per year		22

[1] Filling up fermentor with medium and inoculum.

[2] Discharge fermentor, cleaning equipment, drying the left product

[3] Whole make span. The cycle time is 362.2 hours.

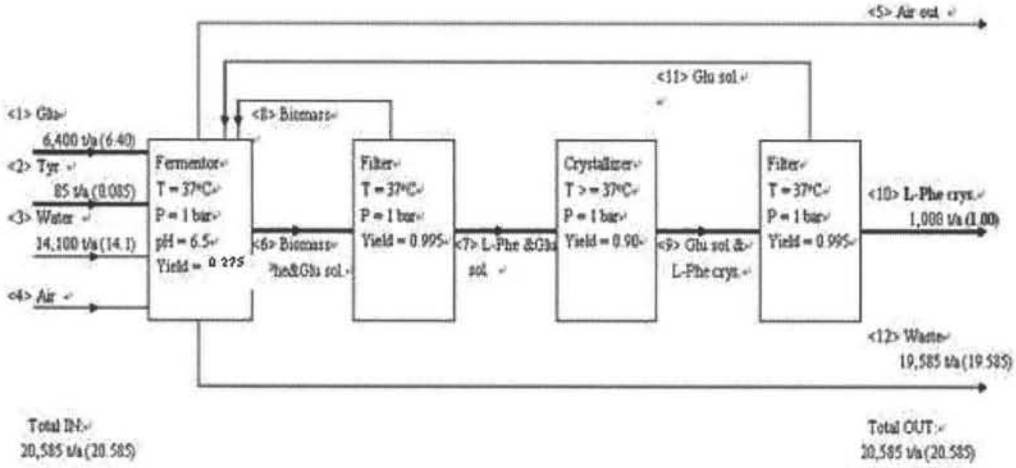


Figure 3.1 Block scheme for the production of L-Phe with a broth concentration of 20 g/l

Integrated fermentation-crystallization ISPR process

After selecting the scheme to develop, several considerations were taken in to account to improve the process.

- Process wastes

  The wastes produced with the scheme selected are very high with high looses of raw materials and the target product.
- The production of secondary compounds must be taken in consideration due to the effect they might have on the process and the product, especially tyrosine that precipitates at very low concentrations.
- The restriction to the type of crystallization method used because of the special properties of the product.
- The time for the duration of each batch cycle (Appendix 5)

The block diagram for improved process to design can be seen in Figure 3.2. Because of the lack of information of the life cycle of the microorganism and the normal time of operation processes in biotechnology the production phase is set to take 14 days[3], the batch cycle time for the process is shown in Appendix 5.

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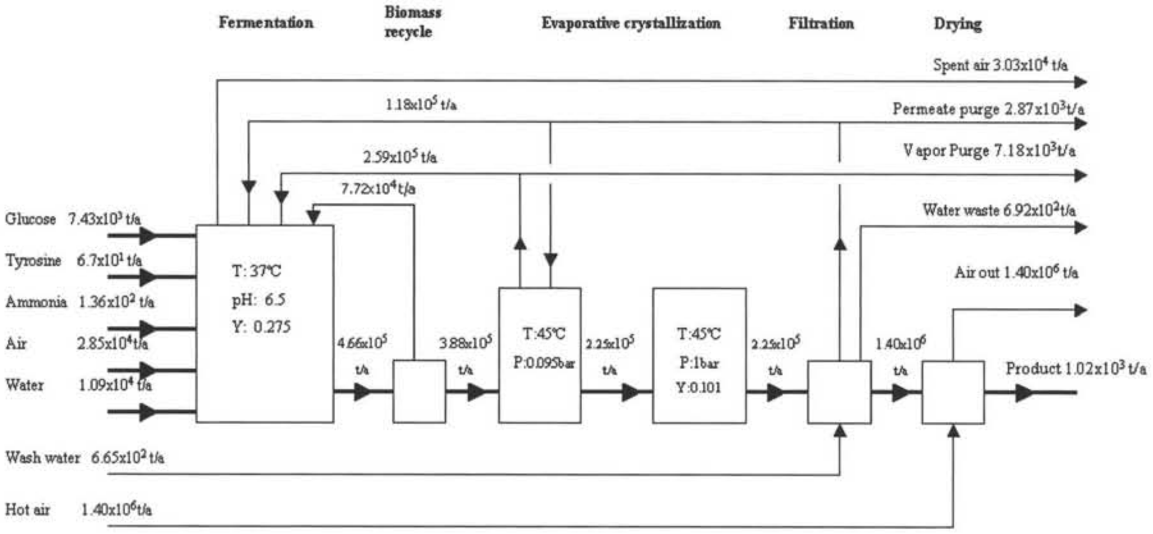
[3] Source: Technical supervisor



Total in 1.44x10^6 t/a **Total in 1.44x10^6 t/a**

Figure 3.2. Block diagram for the design process

In table 3.2 streams entering and leaving per ton of product are listed:

Table 3.2. Streams entering and leaving

Stream	Ton/Ton product
Glucose	7.43
Tyrosine	0.0675
Ammonia	0.136
Air	28.54
Water	10.93
Wash water	0.692
Hot Air	1400
Spent air	30.3
Permeate purge	2.87
Vapor purge	7.18
Waste water	0.692
Air out	1400
Product	1

3.2.3. Thermodynamic Properties and Reaction Kinetics

Little information is available for some of the compounds used in this process, especially for the L-Phenylalanine and Tyrosine, therefore these properties had to be determined using different thermodynamic models described in Appendix 4. For the same reason the values found couldn't be corroborated and assumed to be correct. Other properties for more common compounds like water ammonia, oxygen, amongst others, were taken from literature.

For the development of this process some important factors have to be taken into consideration:

Recombinant strain

Fermentation is done using a modified strain of E. coli that has altered the aromatic amino acids biosynthetic pathway that produces chorismate, this compound is converted trough the three terminal pathways in to L-Tyrosine, L-Phenylalanine and L-Tryptophan. Thus, the production of L-Tyr is suppressed in order to alter the central carbon metabolism and divert the pathway to produce more L-Phe.

L-Tyr plays a key role in signal transduction and in the synthesis of proteins. The microorganism is auxotropic and has to be supplied with sufficient amount of L-Tyrosine in order to grow and survive. The level of L-Tyr has to be controlled carefully especially in the production phase to maintain a constant level of biomass. Furthermore high concentrations of L-Tyr and L-Phe may cause feedback inhibition of the DAHP synthase (Bongaerts *et al*, 2001), a crucial enzyme in the aromatic amino acid pathway. As consequence the central carbon metabolism may be affected and L-Phenylalanine production may be reduced or not be produced at all. For detailed information see Appendix 4.2.

Crystal Polymorphism

An important property that has to be considered is the polymorphism of the L-Phenylalanine, different forms of the crystals show a wide range of physical and chemical properties, including different melting points, structural properties, heat capacities and densities (Mohan *et al*, 2001). L-Phenylalanine can crystallize in two forms, anhydrous and monohydrate. The monohydrate has a needle shape that difficult the utilization and commercialization of the product because of its characteristics, on the other hand the anhydrous type crystals are commercially attractive with better handling and utilization properties.

In order to control the polymorphism of the L-Phenylalanine one has to take in consideration the metastability line with caution of maintaining the product concentration in the solution between this and the solubility line in order to produce the anhydrous crystals. Besides, seeding of anhydrate crystals will accelerate the production speed in the crystallizer and will promote the formation of the required crystal type.

The metastability and solubility lines are shown in Appendix 4.3. The data was taken from the studies made by Mohan et.al, 2001. For detailed information see Appendix 4.3.

3.2.4. List of Pure Component Properties

In table 3.3 properties for the components are listed. The thermodynamic properties for the components involved in the process are listed in Chapter 4.1.1 and can be seen in Table 3.3. The equations for the calculation of these properties can be seen in Appendix 4.1.

Table 3. 3 Lists of Pure Component Properties

PURE COMPONENT PROPERTIES									
Component Name		**Technological Data**					**Medical Data**		**NOTE**
Design	**Systematic**	**Formula**	**MW g/mol**	**Bp °C (1)**	**Mp °C (1)**	**Density kg/m3**	**MAC mg/m3**	**LD50 g**	
L-Phenylalanine	L-Phenylalanine	C9H11O2N	165.19	376.45	172.82	n.a	n.a	220	(2)
Tyrosine	Tyrosine	$C_9H_{11}NO_3$	181.1908	456.85	283.85	n.a	n.a	n.a	
Glucose	Glucose	C6H12O6	180.16	571.29	169.83	1540	n.a	25.8	
Acetic acid	Acetic acid	$C_2H_4O_2$	60.0524	118.1	16.5	1049	n.a	4.96	(3)
Water	Water	H2O	18	0	100	1000	n.a	n.a	
Ammonia	Ammonia	NH3	17	-33	-78	682	n.a	0.35	(4)
Oxygen	Oxygen	O2	32	-182.96	-218.5	1.429	n.a	100 pph/14H	(5), (6)
Nitrogen	Nitrogen	N2	28	n.a	-210	1.2506	n.a	n.a	
Carbon dioxide	Carbon dioxide	CO2	44	n.a	-78.5	1.53	n.a	10 pph/min	(6)
Notes	(1) at 1 bar (2) TDLo oral rat (3) (a) TA (4) (D) at BP (5) (D) at gas (6) (TCLo) inhale								

3.3. Basic Assumptions

Some information needed for the design of the process is not available, most of the information used to design the process is based on lab and pilot scale studies so some assumptions had to be made, actually there is no information for the industrial production of the target compound with the method proposed. The assumptions are made with agreement of the technical and it will be stated otherwise.

3.3.1. Plant capacity

The L-Phenylalanine producing plant design has a product capacity of 1000 t/a distributed in 22 batches per year. The plant life of the designed plant is set to be 15 years and it will operate during the entire year (8000 hr). The operating spam of one batch is 408.7 hours and the production phase spam is 362.2 hours. The product to be obtained is L-Phe crystal in its anhydrous form, because the monohydrate form of the crystal is needle type and is hard to handle so it should be avoided. The crystal mean size is 100 µm and 98.5% purity as desired by the client. L- Phenylalanine is produced from glucose in a fed-batch fermentation process, while tyrosine is the limiting substrate in the production phase because of the use of an auxotrophic L-tyrosine (L-Tyr) recombinant strain of *Escherichia coli*.

3.3.2. Location

The location of the plant is very important for several reasons. To determine the best option to construct the plant aspects as transportation, availability of raw materials and utilities, legislation and environmental considerations have to be made. From an economical point of view the location of the plant has a major impact on the costs of the process such as construction, labor, transport of the product and raw materials as well as the technology and equipments required for the process. The production plant for the designed process is set to be in the Netherlands after consulting with the technical supervisor and considered that the factors stated above are favorable.

3.3.3. Feedstock

The materials needed for the production process are glucose, tyrosine, ammonia and air. The fermentation process uses a recombinant strain of an *E. coli*. Materials such as antifoaming agents and minerals are not taken in consideration.

3.3.4. Battery Limit

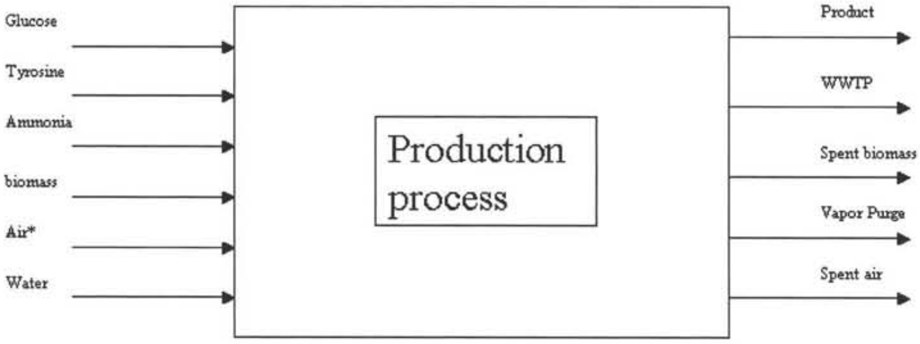
Inside the battery limit the production process is considered and is composed by fermentation and downstream processing. The pre-inoculum is prepared inside the battery limit as well as the raw material solutions with their corresponding storage vessels. The downstream processing is composed of the filtration/washing unit and the dryer. The succeeding required final processes after the drying are not mentioned. The utilities used for the process are assumed to be available from installations close to the plant. The battery limit for the design process can be seen in Figure 3.3 and the units in/out the battery limits in Table 3.4. In Appendix 1 the fractions of the raw materials for batch and extended growth phase are listed.

The feedstock to make the required solution and the biomass for the inoculum used in the process will be acquired outside the battery limits and prepared inside it. In the present report the solutions are considered to be prepared but the time and work load required to prepare the solutions is not consider.

Table 3.4 Definition of units' in- and outside the battery limits

Inside battery limits	Outside battery limits
Inoculum	Sterilization
Fermentor	Pre-culture of biomass
Filters	Acquirement of feedstock
Crystallizer	Acquirement of mineralized water
Downstream processing equipment	Acquirement of filtered air
Auxiliary equipment	Waste treatment

Figure 3.3. Battery limit of the plant



The price and conditions of each stream can be seen in Appendix 3.2

3.4. Economic Margin

In this section, economic margin and the maximum allowed investment will be discussed. The Net Cash Flow (NCF) is calculated by subtracting the cost of the raw materials and waste treatment from the sales income of the product. The economic margin is calculated by dividing NCF by the amount of the product. The total Net Present Worth of the project (NPW_{total}) is calculated on the basis of the NCF in year "n" and discounted cash-flow rate of return (DCFRR). All these calculating methods are described in Appendix 11

Assumptions in this economic evaluation are listed below:

- The life of the plant is 15 years.
- The discounted cash-flow rate of return (DCFRR) is 10%.
- The main raw materials are considered such as glucose, tyrosine, ammonia and water, other compounds with a small amount like seeding, anti foaming agent are not included. Due to the unavailable data for the stream flow of air, cost for air is also ruled out.
- The construction time for the plant is considered to be 2 years.

Based on these assumptions, the main results for economic evaluation of the preliminary design are given in the table below:

Table 3.5 Economic Evaluations

	Costs (M€/a)	Sales Income (M€/a)	Net Cash Flow (M€/a)	Margin (€/kg)	NPW_{total} (M€)	Maximum allowed Investment (M€)
Process	2.88	8.19	5.31	5.31	40.4	40.4

In conclusion, the margin of the process is 5.31 €/kg. The total net present worth of the cash flow is 40.4 M€. Maximum allowed investment based on a certain DCFRR is the value at which the cumulative net present worth at the end of the project is zero, which means at the end of the project the investment will be returned. In this project, it is assumed that the plant (buildings, etc.) already exists and can be put into operation since the first year. Therefore, the maximum allowed investment here is equal to the cumulative present worth of the cash flow: 40.4 M€.

4. THERMODYNAMIC PROPERTIES AND REACTION KINETICS

In order to obtain the thermodynamic properties of the components, some were found in literature other were calculated with the different methods. The equations used in these methods can be seen in Appendix 4.1.

4.1. Thermodynamic Properties

For the components involved in the process, the thermodynamic properties are limited, so a calculation method has to be used for the determination of the different properties. The methods used can be seen in Appendix 4.1. The calculated and reported values for the components are listed in Table 4.1 and Table 4.2

Heat capacities

To calculate the regression coefficients for the heat capacity for the L-Phe, L-Tyr and glucose, the Joback group contribution method was used. The formulas for the calculation of the method, and the results for the mentioned compounds can be seen in Appendix 4.1 (Table A4.1 and A4.2). For the components in Table 4.1 and 4.2 different from L-Phenylalanine, Tyrosine and glucose, the heat capacity was found using the regression coefficients found experimentally and listed in Coulson and Richardson's Chemical Engineering Volume 6. The values found are used in formula 1, for liquids, and 2 for gases in order to find the Heat capacity at different temperatures. The operating window of the process, see Chapter 4.2 stays within the ranges for the application of these equations.

$$Cpl = A + BT + C\frac{T^2}{2} + D\frac{T^3}{3} \qquad 4.1$$

$$Cpg = A + BT + C\frac{T^2}{2} + D\frac{T^3}{3} + E\frac{T^4}{4} \qquad 4.2$$

The group contribution and the Joback values for these components can be seen in Appendix 4.

Boiling and freezing points

Joback group contribution method was applied to determine the boiling and freezing point for the components. These values are of importance because these are input values for the determination of the critical properties. The formulas used for the calculation of the boiling and freezing temperature can be seen in Appendix 4.1.

Critical properties

With the use of the boiling temperature the critical properties are found using the Joback group contribution method.

Acentric factor

To calculate the parameter required for the calculation of the acentric factor the Ambrose-Walton corresponding states method was used (Appendix 4.1) to find the reduced temperature functions f(0)and f(1) along with the critical properties. According to the properties of gases and liquids. Bruce E. Poling, (2001) the average absolute deviation of the calculated value to the reported values in appendix A from the same reference a deviation of 2.4% takes place.

Enthalpy and Gibbs energy

With the Joback group contribution method the Enthalpies and Gibbs energy of formation are calculated using the formulas in appendix 4.1

Table 4.1 Thermodynamic properties of the compounds*

Property	**Unit**	**Tyrosine(1)**	**L-Phe(1)**	**Glucose(1)**	**Acetate(2)**
Tb	K	730.22	649.6	844.44	390.67
Tfp	K	557.69	445.97	442.98	16.6
Tc	K	973.5803	885.5276	1033.968	594.4
Pc	bar	47.6939	39.2094	66.3137	57.9
Vc	cm3/mol	444.5	478.5	459.5	171.5
Δ Hfo	*kJ/ mol*	-491.63	-314.32	-1035.02	-435.13
Δ Gfo	*kJ/ mol*	-297.21	-142.59	-793.74	-376.94
Δ Hv	*kJ/ mol*	15079.3	11969.3	24441.3	23.69
Δ Hm	*kJ/ mol*	7547.12	6165.12	4973.12	n.a.
Zc	*[-]*	0.261895	0.25482	0.354442	0.211
w		-1.93031	-1.58917	-3.47574	-1.2113596
A		-38.709	-29.789	28.561	4.84
B		1.0531	0.8985	0.7419	0.2548
C		-0.00087	-0.00062	-0.00046	-0.0001753
D		2.93E-07	1.6E-07	3.03E-07	4.9488E-08
T	K	298.15	298.15	298.15	298.15
Cp	*J/mol/K*	205.4735	187.5606	216.8546	66.5372119

(1) Thermodynamic properties, The properties of gases and liquids. Bruce E. Poling, (2001).

(2) Coulson & Richardson's Chemical engineering 6th edition. Volume 6

Table 4.2 Thermodynamic properties of the compounds**

Property	Unit	Water(2)	NH3(2)	CO2(2)	O2(2)	N2(2)
Tb	K	373	239.72	194.7	90.17	77.35
Tfp	K	273	195.41	216.58	54.36	63.15
Tc	K	647.3	405.65	304.19	154.58	126.1
Pc	bar	220.5	112.78	73.82	50.43	33.94
Vc	cm3/mol	0.056	72.5	94	73.4	90.1
Δ Hfo	*kJ/ mol*	-242	-45.72	-393.5	0	0
Δ Gfo	*kJ/ mol*	-228.77	-16.16	-394.4	0	0
Δ Hv	*kJ/ mol*	40.68	23.362	17.166	6.742731	5.68225899
Δ Hm	*kJ/ mol*	n.a	n.a	n.a	n.a	n.a
Zc	*[-]*	0.229	0.242	0.275	0.288	0.289
w		0.345	0.252	0.228	0.022	0.04
A		32.243	27.315	19.795	28.106	31.15
B		0.001924	0.023831	0.073436	-3.68E-06	-0.01357
C		1.06E-05	1.71E-05	-5.6E-05	1.75E-05	2.6796E-05
D		-3.6E-09	-1.19E-08	1.72E-08	-1.07E-08	-1.168E-08
T	K	298.15	298.15	298.15	298.15	298.15
Cp	*J/mol/K*	33.65954	35.62391	37.16475	29.37463	29.1765309

(1) Thermodynamic properties, The properties of gases and liquids. Bruce E. Poling, (2001).

(2) Coulson & Richardson's Chemical engineering 6th edition. Volume 6

4.2. Reaction Kinetics

The principal provided the reaction kinetics for this process, they are based on the studies made by Gerigk *et al*, 2002 and Takors *et al.* 2004. The kinetics are specific for the microorganism used, see Appendix 4, which is a recombinant strain of E. coli. The equations are as follows:

$\mu = \mu_{\max} \cdot \frac{C_{Tyr}}{C_{Tyr} + K_{STyr}}$	Growth rate	[1/hr]
$q_{Phe} = q_{Phe\max} \cdot \frac{K_{IPhe}}{K_{IPhe} + C_{Phe}}$	Production rate	[gPhe/gX/hr]
$q_{Tyr} = q_{mTyr} + \frac{\mu}{Y_{XTyr}}$	Tyrosine consumption rate	[gTyr/gX/hr]
$q_{Glu} = q_{mGlu} + \frac{\mu}{Y_{XGlu}} + \frac{q_{Phe}}{Y_{PheGlu}}$	Glucose consumption rate	[gGlu/gX/hr]

With the information provided by Gerigk *et al*, 2002 and Takors *et al*, 2004, the kinetic parameters were determined and are listed in Table 4.3

Table 4.3 Kinetic parameters for the production of L-Phenylalanine

Constant	Value	Unit
μ_{max}	0.321	1/h
K_{STyr}	0.00152	gTyr/L
q_{Phemax}	0.0756	gPhe/gX/h
K_{Iphe}	20	gPhe/L
q_{mTyr}	0.00245	gTyr/gX/h
$Y_{Xtyr,max}$	35.29	gX/gTyr
q_{mGlu}	0.05	gGlu/gX/h
$Y_{XGlu,max}$	0.51	gX/gGlu
$Y_{PheGlu,max}$	0.275	gPhe/gGlu

4.3. Operating Window

In the process designed the operating conditions are mild. Extreme conditions must be avoided in order to maintain a good production without damaging the microorganism or the product itself. The operating window of the process is shown in table 4.4.

Table 4.4 Operating window of the process

Operating condition	Range
Temperature	25-45°C
Pressure	0.095-2.533 bar

4.4. Data Validation

For some of the compounds involved in the process, the thermodynamic data is very limited. Hence in order to check the validity of the method for the calculations of these properties a comparison of the calculated values for acetate against the available data found in Coulson & Richardson's Chemical Engineering 6th Edition and Introduction to Chemical Engineering Thermodynamics 6th Edition by Smith, Van Ness 2001 was made. The reported values for water, oxygen, ammonia, acetate, nitrogen and carbon dioxide were taken from literature. With respect acetate the values were very close to the ones calculated with the Joback method, the errors for this compound ranged between 0.8 to 10%. So for the components whose properties are not found in the literature were found using that method.

5. PROCESS STRUCTURE AND DESCRIPTION

In this chapter the reader will find the information about the process structure of the design. First the design criteria will be explained. Then the selection procedure of each operation unit will be mentioned. The description of the process flow schemes and the utility requirements are explained. At the end the information about the process yields can be also found.

5.1. Criteria

The main design criteria are the purity and the production scale. The desired purity of the produced crystals is over 98.5%. The planned production rate per year is 1000 ton.

Working hours in a year are set to be 8000 hr (Sinnott, R.K, Coulson and Richardson's Chemical Engineering, Volume 6). The process consists of two parts: Batch Part and Production Part. Batch part includes the batch, extended growth, and filling-up phases. The explanation and results of the concentration and mass values for each component are available in Appendices 1 and 7. This study mainly focuses on the production phase. Filling-up phase is neglected because of its minor effect on the process time calculations and it is not taken into consideration in this design study.

After consulting with the technical supervisor, the production phase is taken as 2 weeks. Including the time for batch part, the total batch time is found to be 363 hours. Thus, the annual production rate is achieved making 22.14 batches per year; so in order to fulfill the production capacity 22 batches are planned for the annual production with a product stream of 135 kg/hr. With 1000-ton annual production rate, the production per batch is calculated to be 45.36 ton/batch.

Table 5.1 Production of Phenylalanine

Annual Production Rate	1000 ton
Batches Per Year	22
Batch Time	363 hrs
Production of Crystals Per Year	135 kg/hr
Production per Batch	45.36 ton/batch

A selection of the equipments is done according to their ability to fulfill the set requirements. The equipments are evaluated and selected with the criteria mentioned in Chapter 2 Process Options and Selection. The same criteria and impact factors are applicable for the selection of the equipment.

5.2. Selection

In this part the decision making process during the equipment selection is explained in steps.

5.2.1. Fermentor

A stirred tank fermentor is a widely used type of fermentor. The kinetics associated with the fermentor were given by the client and taken from the same fermentation procedures made by Gerigk *et al*, 2002, Maass *et al*, 2002. Therefore for this design study a stirred tank fermentor is chosen. It was designed by scaling up a 300-L pilot plant fermentor linearly as described in the studies of Gerigk *et al* The medium feeding procedures, oxygen amount needed, and L-Phe production in the fermentation process are determined using these kinetics.

5.2.2. Centrifugation

For the design of the biomass recycle different options were considered and the criterion for the decision was the need for continuous operation and the cost of the equipment. For our process, the biomass has to be removed continuously from the fermentation broth and recycled back to the fermentor. With the use of graph 10.10 from Coulson & Richardson's page 407, the solid liquid separation technique useful for our case considering the percentage of solids in the liquid stream was chosen and two options were feasible: filtration and centrifugation.

For continuous filtration the use of cross flow hollow fiber cartridges was considered, taking as basis, the CFP-2-E-154M (0.2μm) model developed and used by GE Healthcare for the study of the filtration properties of *Pichia pastoris*. This option was rejected because high area will be needed, requiring around 20 filters, 10 filters working at a time, with a high cost, 600000€ (see Appendix 12.1) (Fig 11.8, Bioseparations Science and Engineering) only for the hardware without taking in consideration the membranes. Other types of filters are suitable for the stream characteristics but for batch processes.

To select the type of centrifuge the parameters considered were the volumetric flow, solids content and the continuous separation needed. The best option is a disk centrifuge, with nozzle discharge, according to table 18-12 and 18-13 in Perry's Chemical Engineer's Handbook. In comparison with the previous alternative, the centrifuge is less expensive and has a good performance a seen in Table 2 and 4 from the bioseparations course CE3191 handouts.

The output of the centrifuge was set to 18% (w/w) in order to avoid high concentrations of biomass in the recycle, thus avoiding oxygen depletion in the recycle stream and acetate formation.

5.2.3. Compressor

For the selection of the compressor the parameters to consider are the air flow rate, head of pressure, temperature limitation and power consumption. Using table 10.16 from Coulson & Richardson's page 476, the most suitable equipment for the compression of gas is selected depending on the increase in the pressure and the airflow rate. The appropriate equipment is a displacement rootes compressor and multiple stages to elevate the pressure to the value needed.

Taking in account the discharge temperature, the compressor must operate isothermically and in separate stages with intercooling to avoid high discharge temperatures (387.375 K) that might affect the fermentor performance if an adiabatic compressor were used.

5.2.4. Evaporatorative Crystallization

For the crystallization process, forced evaporative crystallization is selected. Due to the target compound solubility, even at 0°C, the melt crystallization is not applicable in the crystallization process of L-Phenylalanine. On the other hand the polymorphism does not allow working at temperatures below 37°C, because at that conditions the undesired monohydrate crystals are thermodynamically stable. Therefore for the production of L-Phenylalanine a different method must be preferred.

When the solubility line of L-Phenylalanine is observed (Appendix 4.4), the slope of the solubility curve with respect to temperature makes the cooling methodology inefficient. Therefore evaporative crystallization is selected for the crystallization process.

In this study the units composing the evaporative crystallizer - evaporator and crystallizer – are designed as two separate units. This approach gives the following advantages:

- In spite of overall heat requirement of the crystallization equipment, heat duties of each unit have been calculated separately. More accurate estimation for energy duty of the evaporator is reached. Therefore the vessel cooling duty due to crystallization heat is taken into account.
- Secondly and more importantly, the operation conditions can be set differently for each of the two equipments.

The latter point plays an important role especially in economical and safety concerns. The operation temperature of evaporative crystallizer is 45°C and 0.095 bar. Therefore evaporation should take place in vacuum conditions. If the crystallizer unit is designed as one unit, the operation condition of whole unit will be the same.

In the contrary, the separate, or compartmental design of the equipment allows the crystallization operation at atmospheric pressure with the use of a jacket vessel. In this way the high cost of vacuum will be reduced and safety control of the process will be easier.

5.2.4.1.Evaporator

Long tube, vertical forced circulation evaporator is selected for the process. The obligation of the evaporator is to evaporate the required amount of water to obtain supersaturated flow stream. During the decision making process the following factors are taking into account in addition to the process selection criteria mentioned in the previous part (Coulson and Richardson's Chemical Engineering, Volume 6, Chapter 12):

- The nature of the process fluid, particularly its viscosity and propensity to fouling

The process liquid is dilute. Having almost the same properties with water viscosity and fouling is not a problem. Therefore nature of the process fluid does not play a significant role in the decision process.

- The operating pressure: vacuum or high pressure.

In contrary to the first one, the second criterion is an important factor determining the type of the evaporator to be used. Because of having proteins in the liquid medium, high temperatures are avoided for the evaporation. High temperature may lead denaturalization of the protein. Consequently vacuum conditions should be sustained to evaporate the water at low temperatures, however, vacuum is usually not a desired operation due to the high operation costs and safety concerns. The operation pressure is set to be 0.095 bar, which is the boiling point of water at the operation temperature of 45°C. The reasoning of the temperature value is explained in the next page, while the effect of solubility criterion in the selection of type of evaporator.

The requirement of high vacuum operation makes the type of forced circulation evaporator favorable for the process despite the high operating cost due to pumping cost.

- The equipment layout, particularly the headroom available.

Horizontal exchangers are more advantageous than the vertical one due to their less headroom requirement. However when the types of evaporation working with natural convection are considered, the high heat duty requires very large surface area. To decrease this demand for high area, the overall heat transfer coefficient must have a high value.

"The liquid-film transfer coefficient can be increased by pumping to cause forced circulation of the liquid. This could be done with the use of long-tube vertical type by adding a pipe connection with a pump between the outlet concentrate and the feed line." (Geankoplis, C.J., *Transport Processes and Unit Operations*, Third Edition, Prentice Hall, India, 2000)

According to the same source, the overall heat transfer coefficient of forced circulation evaporator systems for long-tube vertical applications varies between 2300-11000 W/m^2*K which is very high in comparison with the natural circulation, 1100-2800 W/m^2*K.

- Solubility of the component

High risk for crystallization of phenylalanine inside the evaporator and its low solubility would be problem in a system where the load of evaporation is as high as in this process; 54.2 w% of the inlet stream is evaporated.

The low solubility of components and local temperature differences usually result in crystal formation inside the evaporator. High velocity rate of forced circulation will sustain higher homogeneity inside the liquid medium preventing the crystal formation. Also because of this specificity the use of forced circulation is more advantageous for this process in comparison with its alternatives, which are open kettle or pan, horizontal-tube natural circulation evaporator, vertical-type natural circulation evaporator, long-tube vertical-type evaporator, falling-film type evaporator, agitated-film evaporator and open-pan solar evaporator.

Determination of temperature leaving out the evaporator is determined according to the properties of phenylalanine crystal formation. There were some constraints to be taken in account and that the system should obey. The first constraint limiting the value of temperature is its denaturalization temperature. The resistance of the phenylalanine against denaturalization is not known exactly, but after the meetings with the supervisor it is concluded that Phenylalanine can resist the temperatures up to around 60°C. In order to minimize the risk of denaturalization, the operating temperature should be set less than this limit.

The other constraint limiting the range of temperature is the solubility curve of the product L-Phenylalanine. Since the stream will be crystallized inside the crystallization vessel after the evaporator, the condition should make the crystallization of anhydrate crystals thermodynamically favorable, condition obtained at temperatures higher than 37°C.

Although, according to the solubility curve at higher temperatures higher yield of crystal formation is obtained, the change in yield is very little; and furthermore high operating temperatures would result in higher duty of evaporation the energy cost would be higher. Taking into account this concerns the temperature of the evaporator is set to be 45°C and supersaturated phenylalanine concentration is set as 44 g/L, where the metastability zone ends at around 44.9 g/L.

In summary, design operation conditions for evaporation are 45°C, and 0.95 bar vacuum pressure.

5.2.4.2. Crystallizer

In order to select the crystallizer type, the properties of the target component were evaluated. The selected process was to use an evaporative crystallizer; the cooling crystallization and melt crystallization where rejected because of the low concentration leaving the fermentor, the type of crystal required and the solubility properties of the L-Phenylalanine, see Appendix 4.

In order to produce the crystals needed for each batch, the volume of water to be evaporated is very high and due to previously mentioned reasons it was split in to an evaporator and a crystallizer with the same function as en evaporative crystallizer where the water is heated in the forced circulation evaporator and the crystallization takes place in the crystallizer at the supersaturation level without a change in temperature. A vacuum evaporator is used to concentrate the product at relatively low temperature without damaging the amino acid, and achieve the supersaturation in the crystallizer required to meet the annual production.

5.2.5. Frame Plate Filter

The plate and frame press filter is used to separate L-Phe crystals in this process. In our process, the feed stream rate is very high (about $30m^3/h$) and crystals in it are only 0.45% by mass. The formation of cake will increase the total resistance. So, in order to obtain a high overall filtration rate, higher pressures in equipment are required according to Darcy's law (Coulson, J. M, Richardson J.F.; Coulson and Richardson's Chemical Engineering, Vol. 2, Edition 4,p.132). Also, the cost of the equipment will be closely related to the filtering area, which can be less if the flow rate is high. Therefore the plate and frame press is chosen by its simplicity, low capital cost, and ability to operate at high pressure.

The rotary drum filter was evaluated and rejected because of the low-pressure differentials and as consequence, large area is needed. In order to form enough thickness cake, the rotary rate is very slow, around several hours, and according to Perry (Chemical Engineering Handbook) the cycle time range for rotary filters is around 0.1 to 1 r.p.m.
The centrifuge is not efficient equipment in our process either. The cost is very high as well as the cost for its operation. Furthermore another unit in the process would be needed for the washing of the crystals.

Another option for the filtration of crystals was considered and represented in Appendix 12.2, hydro-cyclone rotary drum filter. This option did not chosen as the final one because of the lack of information about the use of this equipment.

5.2.6. Dryer

Because L-Phenylalanine is a heat sensitive compound, the best way to dry it is at low temperatures. According to Table 16.7 (Coulson, J. M, Richardson J.F.; Coulson and Richardson's Chemical Engineering, Vol. 2, Edition 4,p.731), a pneumatic dryer is selected to dry the crystals because the rate of drying is fast, thus protecting the product from prolonged periods of drying.

5.2.7. Condenser

The vapor leaving out the evaporator is at vacuum pressure, 0.095 bar, and has to be condensed. Condensation is done using cooling water. Two types of condensers can be used for this purpose: surface condenser and direct-contact condenser.

Although at the first look direct-contact condenser, which allows higher overall transfer coefficient, would seem be the best solution due to having water in the both sides, there are specific problems inhibiting the use of direct-contact condenser:

Vapor leaving the evaporator is saturated steam at 0.095 bar and 45°C. For the use of a direct-contact condenser the steam must have higher pressures. The pressure would be increased by the use of compressor. However the change in pressure of the steam would immediately result in phase change inside and the presence of water inside the compressor would certainly hinder its operation.

Because of these concerns, surface condenser is chosen as the type of condenser. It has exactly the same mechanism with a heat exchanger.

Although the evaporated steam with load of 35760 kg/hr has a large energy capacity, its low temperature prevents the use of this energy to heat the process liquid in another part of the process. In process steam is used only in the evaporator to evaporate water at 45°C and because of having the same temperature the steam cannot be used in the evaporator to evaporate the water, but it can be used to heat the process flow <120> coming into the evaporator from 37°C to around 45°C where the further heat transfer is prevented thermodynamically due to small temperature differences between two streams.

5.3. Process Flow Scheme

There are two different types of flow sheets in appendices: process flow scheme describing the continuous production phase (Appendix 5.1.1), and batch flow scheme describing the batch procedure (Appendix 5.1.2).

5.3.1. Description of Continuous Process Flow Scheme

The raw materials of the process are glucose, tyrosine, and ammonia. They are introduced to the fermentor (R101) via pumps (P101), (P102), (P103), respectively. Air at 25oC and 2.5 bar enters to the fermentor after being compressed (K101). The concentration of air inside the fermentor must be controlled and kept around 40% dissolved oxygen according to Rüffer *et al*, 2004 in order to inhibit the anaerobic acetate production. With this purpose a flow control unit is adjusted to the air inlet flow.

The operating conditions of fermentor (R101) are 37°C and 1 atm. The temperature level is kept constant by the use of coils where steam rate is controlled by temperature controller. Being open to the atmosphere with the air outlet flow stream the pressure does not need to be controlled with an additional control unit. The detailed information about the process control of the system can be found in the next chapter (Chapter 6).

The fermentor broth liquid is pumped continuously via pump (P105) to the centrifuge (S101). The function of the centrifuge is biomass removal. The additional vessel unit, (V105), is set due to control reasons. The removed biomass is recycled back to the fermentor via line <116>. The effluent, <118>, is pumped to the evaporator. The mass fraction of biomass in line <116> is set to be 18% to avoid oxygen depletion in the recycle and as consequence anaerobic formation of acetate as mention before.

The operating conditions of evaporator (V101) are 45°C and 0.95 bar. The concentration of phenylalanine is increased in this unit to the supersaturation concentration 44 g/L from the value in the fermentor, 15 g/L. The vacuum operation requires good control. Fluctuation in the operation conditions and inhomogeneous concentration or temperature dispersion would result crystal formation inside the evaporator, which is not desired.

The steam evaporated <135> is condensed in compressor (E102) at 0.095 bar. The condensed water stream <138> is splitted into two streams, <139> and <140>. Stream <140> is the purged water stream and <139> is the water returned back to the fermentor (R101).

The process liquid leaving out the evaporator (V101) is pumped to the crystallizer (V102). The temperature in crystallizer is 45°C and solubility of phenylalanine at that temperature is 39.73 g/L. The desired form of the crystals is anhydrous and due to having a continuous operation seeding is required only at the start of run.

The stream leaving the crystallizer, <121>, is pumped via pump (P108) to the plate frame filter, (S102), for the separation of the crystals. Type of operation of the filter is batch; therefore two filters are used in parallel, where one is used as the other one is cleaned. The produced crystals are washed with water stream <145> coming in to the system via a pump (P111).

The effluent stream <130> having dissolved phenylalanine composition 39.73 g/L is splitted into three streams. Stream <132> introduced to the evaporator for the recovery of dissolved phenylalanine; Stream <133> returned to the fermentor (V101); and Stream <134> is the purge stream set to inhibit accumulation in the process.

The crystals with water content of 37.5%, without taking in consideration the mass of the dissolved components (2.5%), are removed from the plate frame filter (S102) with line <125> and transported to the dryer (D101) by the use of a conveyor belt (X101). Hot air, <127>, is used to dry the crystals. The product stream, dry crystals leaving out the dryer at the bottom, is shown with stream number <128>.

5.3.2. Description of Batch Process Flow Scheme

Fermentation

Glucose and tyrosine solutions are fed with water to the fermentor via pump (P101), pump (P102) and pump (P104) respectively, the feeding time span is around 0.7 hours. After that, the inoculum solution is added to the fermentor for approximately 0.3 hours.

Subsequently, the aeration (K101), pH control (P103), temperature control and off gas condensation (E101) is started at the beginning of the batch phase. The batch phase will take 9 hours. When the concentration of glucose and the concentration of tyrosine are lower than 5 g/L and 0.01 g/L respectively. The feeding of the Glucose and tyrosine are made using the same pumps as before until the end of the production phase.

The operation of in situ product removal begins at the point when the L-Phenylalanine concentration in the fermentor reaches 15g/L (around 22.3 hour) and it last until the end of the production phase (358.3 hour). In this period, most of the equipments are used and operated continuously, including fermentor (R101), centrifuge (S101), evaporator (V101), crystallizer (V102), filter (S102 A/B), dryer (D101), condensers (E101 and E102), compressors (K101, K102 and K103), and pumps (P101 to P111).

After 358.2 hour corresponding to the end of the Production phase, the aeration (K101), glucose (P101), tyrosine (P102) and ammonia (P103) feeding is stopped. Then, the fermentor discharge and equipments cleaning takes 1.9 hours and 2.0 hours respectively. In the end, the last batch of the crystal from filtration is dried in the dryer (D101) using 48.5 hours.

Filtration and Drying

Since the plate and frame press filter is batch filtration, two filters (S102 A/B) are used alternatively to ensure the continuous process in the production phase. The filtration time for one of the filter is 48 hours, after this, he crystal washing and discharge will take about 0.6 hours. When filter A stops filtration, filter B continues with the filtration, while, at the same time, crystals in filter A are washed and discharged. So as a conclusion, 7 batches of filtration are enough for one batch cycle using the switch operation mode described above.

Dryer is started at the first batch of crystals discharged from the filter (70.9 hours), and stopped till the last batch of crystals is dried (408.7hours).

5.4. Batch Cycle Diagram

Table 5.2 shows the typical time of the L-Phenylalanine production process in one batch. The specification operation steps and the origin of these times are shown in Appendix 5. Figure A5.1 (Appendix 5) describes the number of batches performed in one year. In order to assure the required production capacity (1000t/a), 22 batches are needed in one year. The details of the batch time are given in Figure A5.2 and Table A5.2. We can obtain that the make span of one batch is 408.7 hours, while the cycle time of the process is 362.2 hours, this difference is seen because the dryer can be still performing when the new fermentation cycle is started.

Table 5.2. Operation times of processes

Process	Time	Unit
Starting up[1]	1.0	[h]
Biomass growth in batch phase	9.0	[h]
Extend growth phase to steady state production phase	12.3	[h]
Steady state production phase	336.0	[h]
Discharging fermentor	1.9	[h]
Cleaning	2.0	[h]
Drying [2]	48.5	[h]
Total make span	362.2	[h]
Total cycle time	408.7	[h]

[1] Include feeding every medium and starting biomass

[2] Drying time needed after other operations finished.

5.5. Process Stream Summary

In Appendix 5 mass flows are tabulated for all the components separately and in total for each stream. Temperature, pressure, phase and enthalpy values of each stream are specified. These tables can be used to check the accuracy of the mass and energy balances.

5.6. Utilities

Fermentor

The cooling duty in the fermentor is 2314 kW, and the required amount of cooling water is 91,740 kg/hr. Water is coming in at 10oC and going out at 30oC. On the other hand the power input of the stirrer is 232 kW. The detailed calculations about the utility can be found in CD (fermentordesignfinal.mcd).

Evaporator

For a heating duty of 24,270 kW condensed steam at 100°C is used and it condenses totally at the same temperature during the heat transfer. The amount of steam required is

38710 kg/hr. Calculations related with the utility consumption of evaporator can be found in Appendix.8 Equipment Design.

Condenser

For the cooling duty of 24,120 kW, the entrance temperature of the cooling water is 5°C and the outlet temperature is 25°C. The hourly water requirement is 1,038,000 kg/hr. Calculations related with the utility consumption of condenser can be found in Appendix.8 Equipment Design.

Crystallizer

Cooling duty of crystallizer is 337.315 W. This little amount of heat duty can be obtained by the use of a heating jacket. The cooling water is coming in at 10°C and leaving the crystallizer at 25°C. The amount of need for water can be calculated then as:

$$\text{Amount of water} \quad Q / Cp * \Delta T = 377.442 \text{ kg / hr}$$

The available area of the heat jacket is equal to the area of crystallizer. But is this area enough for the heat transfer of required amount of heat? The height of the crystallizer is 1.526 m and the diameter is 1.017 m. Overall heat transfer coefficient of cooling water is in a range of 3000-6000 $W/m^2 * {}^\circ C$. It is assumed to be minimum, which is 3000 $W/m^2 * {}^\circ C$.

$$\text{Calculated Area} \quad H*D*\pi = 4.877 \text{ m}^2$$

$$\text{Required Area} \quad Q / (U*\Delta T) = 7.5 \times 10^{-3} \text{ m}^2$$

Therefore the heat duty can be satisfied by the use of a jacket efficiently. The summary of utility consumption can be seen in utility Table 5.3:

Table 5.3 Utility Tables

	Heat Duty (kW)	Steam (kg/hr)	Cooling Water (kg/hr)	Inlet Temperature (°C)	Outlet Temperature (°C)
Fermentor	-2314	-	91740	10	30
Evaporator	24270	38710	-	100	100
Crystallizer	-0.337	-	337.31	10	25
Condensor	-24120	-	1.04×10^6	5	25

5.7. Process Yields

The yields of in- and outgoing streams per kg of L-Phenylalanine crystals were calculated. The value of yields and the related stream numbers are stated in Table 5.4.

Table 5.4 Yields in Raw Materials per Product

Component	Stream number	IN [kg/kg]	Stream number	OUT kg/kg	Process kg/kg
L-Phe		1.599E-05	<128><134> <146>	1.000E+00	1.00
Glucose	<101><102>	6.490E+00	<128><134> <146>	4.961E-02	-6.44
Tyrosine	<103><104>	5.895E-02	<128><134> <146>	2.823E-08	-0.06
Ammonia	<105><106>	1.188E-01	<128><134> <146>	4.466E-04	-0.12
Water	<101><102> <103><104> <105><106> <107><108>	1.028E+01	<134><140> <146><129>	9.745E+00	-0.54
Gas	<109><110>	2.494E+01	<113>	2.648E+01	1.54
Hot air	<127>	1.221E+03	<129>	1.222E+03	0.55

Note: The minus sign signifies consumption

6. PROCESS CONTROL

Since the whole process is submitted to a certain set of conditions, control is required by the system in order to maintain these conditions. In this chapter various controller for different uses will be discussed. Only basic control is introduced, advanced control is beyond the scope of this project. For those controllers placed for safety use will be discussed in more details in Chapter 10 the safety section.

The process flow sheet including the controllers is shown in Appendix 5.

6.1. Fermentation Section

Due to the feeding complexity during the fed-batch process, flow control on glucose and tyrosine-feeding stream is required. Glucose (700g/l) and tyrosine (25 g/l) are fed to the fermentor after the initially supplied substrates were consumed (OD_{620} 12-15). The growth is controlled by the addition of limited amounts of L-Tyrosine to the media. When an OD_{620} around 80 is reached the tyrosine supply is reduced to a minimum of 0.01g/l sufficient for maintenance. Glucose is controller at a set point of 5g/l in order to prevent acetate formation or limitations resulting in lower product formation rates. Flow control is placed on glucose, tyrosine and ammonia water feeding stream (streams <102>, <104>, and <106>) and after pump.

Inside the fermentor temperature is controlled at 37°C. This temperature is controlled by a temperature controller, which controls the flow of the cooling agent. If the temperature in the fermentor gets too high, the flow rare of cooling water stream is increased. If the temperature in the fermentor gets too low, the flow rate of cooling water stream is decreased. The pH in the fermentor is kept at 6.5 by placing a pH controller (YC1) on ammonia feeding stream<106>. The air flow through the fermentor is controlled with a set point provided by the dissolved oxygen tension controller YC2. Since the off-gas leaving the fermentor contains a certain amount of vapor, the liquid level in the fermentor will be decreased. The liquid level in the fermentor is controlled by the level controller which manipulates the flow rate of stream <115> leaving the fermentor to the centrifuge.

This controller keeps the level in the vessel as close as possible to the set point, herewith avoiding that the level reaches abnormal higher or lower values. Due to the loss of the vapor in the off-gas, a condenser E 101 is added in order to cooling down the vapor in the off-gas. Therefore a temperature controller manipulating the flow rate of the cooling water is placed on E101.

6.2. Centrifuge

In order to control the flow rate of the stream which leaves centrifuge to the evaporator, an additional storage tank V105 is added after the centrifuge. A level controller manipulating the flow rate of stream <119> is used to control the liquid level in V105.

6.3. Evaporator

Temperature controller is placed on the evaporator in order to control the temperature at 45°C by manipulating the steam flows for heating the evaporator. If the temperature in the evaporator gets too high, the flow rate of steam stream is increased. If the temperature in the evaporator gets too low, the flow rate of steam stream is decreased. The vapor continuously leaving the evaporator will lead to the fluctuation of the liquid level in the evaporator. The liquid level in the evaporator is controlled the level controller manipulating the flow rate of the stream<122> leaving the evaporator to crystallizing tank (V102). Pressure in the evaporator should be kept as 0.095 bar. Since the pressure in stream<119> and <132> is 1.1 bar, a valve is placed on stream <120> in order to keep the pressure in the stream going into the evaporator as same as that in the evaporator. The vapor leaving from the evaporator is condensed through condenser E102 and then recycled to the fermentor. For smooth and steady operation it is necessary that the condensor is controlled. This makes sure the design value of water is condensed from the off gas stream leaving the evaporator (stream<135>). This can be achieved by cooling the off gas to a certain temperature (37°C). The temperature of stream <136> is a measure for the amount of water condensed. Measuring the temperature of stream <136> and adjusting the cooling water flow is a good way of controlling the amount of water condensed from stream <135>. A vessel V104 is added after the condenser E102 to temporarily store the liquid. Hence a level controller is required to maintain the liquid level in the V104 by manipulating the flow rate of the stream leaving V104. In order to keep the pressure in the evaporator as vacuum (0.095 bar), a compressor K102 is necessary to suck the air out. The fluctuation of the flow rate of the air out which influence the pressure in the evaporator will be compensated by another air stream<141> going into V104. Therefore, a pressure controller is required to maintain the pressure in the evaporator as 0.095 bar by manipulating the flow rate of the air stream<141>. In addition, stream <103> after P110 is split into two streams stream <139> and stream<140>. Flow controller is placed on stream <140> to keep a certain fraction of this stream.

6.4. Crystallizer

Temperature controller is placed on the crystallizer to control the temperature in the evaporator as 45°C by manipulating the steam flows for cooling duty of the crystallizer. Cooling jacket is used to keep the temperature inside the crystallizer constant. Due to having a very low cooling duty the control of temperature can be sustained perfectly. Also level controller is necessary to maintain the liquid level in the crystallizer by manipulating the flow rate of the stream leaving the crystallizer to the plate-frame filter.

6.5. Plate-Frame Filter

Due to the pressure drop in the plate-frame filter, the pressure of the stream going into the plate-frame filter should be maintained to a certain value in order to push the permeate out of the filter. With the stream <124> continuously entering into the filter, the thickness of the cake in the filter gradually increased, which result in an increase of pressure drop in the filter. In order to maintain the flow rate of the stream <130> leaving the filter, a pressure controller is required to control the pressure of stream<124> to the set point by manipulating the flow rate of this stream. In the case when the control valve is 100 per cent opened while it still cannot satisfy the requirement of the increased pressure drop, the controller is not reliable at this moment. Therefore, it is necessary to place a pressure drop sensor to measure the pressure drop inside the filter. If the pressure drop exceeds the maximal value, the thick cake has to be removed from the filter. Stream <131> leaving the plate-frame filter is split into three streams (stream<132>, stream<133>, stream<134>). Flow controllers are placed on stream <132> and stream <133> respectively to maintain the fraction of the flow recycled to the evaporator and that of the purge solution.

7. MASS AND HEAT BALANCES

Mass and energy balances are found in Appendix 5 and 7. Balances gave very good results and as a result of detailed calculations it converged with a very high accuracy, 99.8%. The small difference can be explained because the calculations and the stream tables were made using MathCAD and Excel respectively. In this chapter the calculation process is described and the constraints taken into account during the calculations are explained.

7.1. Mass Balances

Due to having four recycle streams in the process the calculations of the system must be solved altogether at the same time. MathCAD program is sufficient to solve this kind of systems in a loop and therefore it has been used in mass balance calculation.

Overall mass balances, component mass balances – phenylalanine, tyrosine, acetate, ammonia, glucose, oxygen and carbon dioxide, and some specifications are solved in one solve block and further simple calculations are done afterwards. The MathCAD file of the calculations can be seen in Appendix 7.

Some constraints settings are specific for each equipment. MathCAD is configured to change the split factor of the recycle streams and purge streams in order to satisfy the constraints given.

Fermentor

Mass balances must satisfy fermentor operation conditions set by the studies of *Takors et. al,* 2004. As mentioned in Chapter 5 the fermentation operations are based on the studies made by Gerigk *et al.*, 2002 and *Takors et. a., 2004* in this way:

- Glucose concentration is kept constant at 5 g/L inside the fermentor.
- Biomass concentration remains constant at 30 g/L (Zero growth rate during the process).

- Tyrosine concentration should be exactly in the amount that is just enough for the maintenance of the biomass and that leads to zero growth rate. Tyrosine having similar thermodynamic properties with phenylalanine is not desired to be present in the crystallizer also the solubility of the tyrosine is very low (0.38 g/L), thus, present in the crystallizer would increase the impurity of the product and the product would require further stages for purification.
- Dissolved oxygen concentration is equal to 40% of its solubility at that temperature and pressure.
- Undesired acetate production should be kept as low as possible. By the approval of team supervisor, the following assumption is made: 0.5% of the glucose entering to the fermentor is consumed for acetate production. High oxygen concentration inside the fermentor is thought to be enough to obtain low production rate of acetate.
- In addition, the concentration of acetate inside the fermentor is set to be less than 5 g/L.
- Ammonia is fed to the system just in the amount consumed and to keep the pH constant at 6.5. Its concentration is set to be 0.1 g/L.
- Phenylalanine concentration is kept at 15 g/L, because at low values of L-Phe the production rate is higher as explained in chapter 3.
- Beside these, dissolved carbon dioxide concentration is set to be equal to its solubility at that conditions, due to high rate of carbon dioxide production in the fermentor.

Evaporator

- Biomass is perfectly removed by the centrifuge. No biomass content is found in the evaporator and in the following units.
- Phenylalanine concentration is 44 g/L, supersaturated concentration.
- Oxygen and carbon dioxide concentrations are set to the solubility values at evaporator conditions.

Crystallizer

- Glucose concentration in the crystallizer must be kept below 16 g/L. High concentration of glucose is not desired for the purity concerns of the formed L-Phenylalanine crystals. With the recommendation of group supervisor glucose concentration is wanted at the first stage below 10 or 15 g/L. Convergence of the mass balances were the problem at the first point. Although the balances converged at around 14.9g/L glucose concentration, due to optimization reasons this concentration was set at 16 g/L. The reason for this configuration is the significant drop in the amount of water to be evaporated (almost 50%) only by a change of 1 g/L in glucose concentration. Therefore glucose concentration is decided to be 16 g/l.
- Dissolved phenylalanine concentration is 39.73 g/L, which is equal to its solubility.

Crystal Filtration

- The dry crystal content of the filtered product stream is set as 60% by weight. During the literature research, crystallization and filtration processes applied in industry are investigated. Patents were searched with this purpose. In the patent "Method of Preparing Alpha-L-Aspartyl-L-Phenylalnine Methyl Ester" with the publication number EP 0514936A1 (Applicant: Ajinomoto Co., INC., year: 1992), wide information on the water content of L-Phenylalanine crystals is found. Dependent on the filtration process applied, water content of crystals varies between 28-72%, but for many applications this percentage is around 38-45%. Therefore 60% dry crystal content (40% liquid broth) is found to be reasonable and is applied in the design.

7.2. Energy Balances

After the overall and component mass balance calculations energy balance calculations have been done to determine the heat duties for the equipments in the process.

The heat capacities of each component at the different temperatures are calculated using the regression constants calculated with Joback method as seen in Appendix 4.1.
The value of the heat capacity for each stream is calculated using the enthalpy of the components at the temperature of the relevant stream. For the Energy balance calculation the reference temperature is taken as 25 ºC.

After the mass fraction calculations for each stream, the stream enthalpies are calculated multiplying them by the enthalpy of the components at the specific temperature of the stream. Thus, the enthalpies of the streams are found and can be seen on Appendix 5, Table 5 .3.

In order to determine the heat duty of equipments, energy balances are done around them. The heat addition or heat removal job is done by utility consumption.

8. PROCESS AND EQUIPMENT DESIGN

8.1. Process Simulation

For modeling the fermentation process and all design calculations are done in MathCAD program. MathCAD was a very useful program to deal with the calculations. Its easy use and the well knowledge of the group members in this program are the main reasons for its selection.

Because of being an integrated process system includes four internal recycle streams which make the calculation of all balances altogether. Including four loops in the mass balances, the successful iteration was requiring high level of convergence. In the calculations all the stream flows are calculated within the same program. The constraints are also set in the same program. Constraints are explained in Chapter 7.1 Mass Balances.

Optimizations have been done according to the criteria set (Chapter 5). Optimization of the process has been done according to utility use, equipment duty change, production increase and yield. The constraints are changed and its effect is investigated. In setting the constraints the studies of Gerick et al., 2002 and Takors et al, 2004 are used. However all the required information was not available in the articles and some decisions must be taken. At this point the supervisor of the team, with his expertise Dr.Ir.Adrie Straathof helped the team in taking decisions. The constraints and their reasoning are described in Chapter7 in detail.

However in the calculations the constraints have been changed sometimes. During the calculations, little changes in constraints have been done for compulsory and economical reasons:

In spite of its effect on L-Phenylalanine crystallization is not known exactly, glucose is not desired to be present with high concentrations inside the crystallizer because of purity reasons. Therefore its concentration is set to be less than 10 g/L at meetings done by the supervisor. However when the balances did not converge, this constraint is changed. The convergence has been satisfied around 14.93 g/L of glucose concentration inside the crystallizer. Hence the constraint changed to 15 g/L.

Afterwards, in the optimization studies it is seen that a change in glucose concentration by 1 g/L decreases the evaporation duty approximately 50% (from around 71000 kg/hr to 35760 kg/hr). Therefore for economical reasons glucose concentration set to 16 g/L.

8.2.Equipment Selection and Design

8.2.1. Fermentor

A stirred tank fermentor is chosen. According to N. Rüffer *et al* 2004, the dissolved oxygen in the fermentor during the production phase should be maintained at 40% in order to avoid the anaerobic fermentation of the glucose and as consequences the production of acetate. The amount of oxygen required by the microorganism and the excess needed to maintain the dissolved oxygen concentration in the liquid was calculated as seen in the mass balances see CD with the MathCAD file (Mass and Energy Balance Claculation.mcd). The results show that the use of air as the only oxygen supplier cannot satisfy this requirement. Therefore, mixing gas with air and pure oxygen was chosen for the aeration of the fermentor. The air stream is pumped to the fermentor using a compressor to ensure the good aeration and the oxygen levels needed for the process. A condenser in the gas exit is needed in order to recover the water vapor taken by the off gas and return the water to the fermentor. The main operating conditions of the fermentor are shown in Table 8.1

Table 8.1 Fermentor operating conditions

Parameter	Value	Unit
Liquid volume	110	[m^3]
Temperature	37	[^{o}C]
pH	6.5	[-]
Pressure at the top	1	[atm]
Fermentation time[4]	357	[hr]
Aeration rate	3.157E+3	[m^3/h]
Cooling water rate	9.174E+4	[kg/h]

8.2.2. Centrifugation

A disk centrifuge, with nozzle discharge, was selected according to table 18-12 and 18-13 in Perry's Chemical Engineer's Handbook see Appendix 8. In comparison with the other alternative, continuous filtration as seen in Chapter 5.2.2, the centritrifuge is less expensive and has a good performance a seen in Table 2 and 4 from the bioseparations course CE3191 handouts.

The output of the centrifuge was set to 18% after a conversation with the technical supervisor in order to avoid high concentrations of biomass in the recycle, thus avoiding oxygen depletion in the recycle stream and acetate formation.

8.2.3. Evaporator

Design operation conditions of forced circulation type evaporator are 45^{o}C, and 0.095 bar vacuum pressure. 8m long vertical tubes are used to evaporate 35760 kg/hr, 54.2 w% of the inlet stream.

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[4] Include batch growth time, extend growth time and production time.

Table 8.2 Design Information of Evaporator

Heat Duty (kW)		24270
Process Flow (kg/hr)		66010
Temperature (^oC)	In	37
	Out	45
Overall Heat Transfer Capacity (W/m*^{o}C)		3589
Area (m^2)		375.895
Number of Tubes		785.11
Tube Diameter (mm)	Inside	14.83
	Outside	19.05

8.2.4. Crystallizer

The design of the evaporative crystallization device is split in two equipments for calculation purposes (Chapter 5.2.4). First the evaporation section of the crystallizer is designed and then the crystallizer. The crystallizer will consist on a tank with a cooling jacket that will remove the energy of solution needed to form the crystals. The energy that needs to be removed is negligible in comparison with the energy needed to evaporate the water.

Therefore a little amount of heat, 337.315 W, has to be taken out from the vessel to keep the temperature constant at 45°C. For detailed information, in Chapter 5.6 utility requirements can be seen.

Size distribution

The average required size of the crystal product is 100μm. Crystals with the size larger than the required size will be removed from the crystallizer, while crystals with the size smaller than the required size will be recycled to the crystallizer. The desired particle size distribution by the client is between 50μm and 150 μm but these calculations are not taken in account in the present work.

8.2.5. Plate-Frame Filter

The plate and frame press filter is used to separate L-Phe crystals in this process. In our process, the feed stream rate is very high (about $30m^3/h$) and crystals in it are only 0.45% by mass.

The range of our product size is between 50 μm to 150 μm, so the woven fabrics with 10 μm are suitable for the filtration. Since the plate and frame press is batch operation, one more of it is needed to do the continuous operation in the whole process. The operation pressure should increase gradually with the increase of cake thickness to maintain the constant permeate flow rate. The operation conditions of the plate and frame press are given in Table 8.3. The washing water per batch is also calculated and the detailed calculations are in Appendix 8.5.

Table 8.3 Design Information of Plate and Frame Press Filtration

Parameter	Value	Unit
Plate size	1.5x1.7	[mxm]
Plate number	6	
Pore size	10	[μm]
Filter area	30.6	[m^2]
Permeate rate	0.92	[$m^3/m^2/h$]
Maximum transmembrane pressure	3.2	[bara]

8.2.6. Dryer

The L-Phe crystal discharged from the filtration contains about 40% water and impurities. Because L-Phe is a heat sensitive compound, the best way to dry it is at low temperatures. According to Table 16.7 (Coulson, J. M, Richardson J.F.; Coulson and Richardson's Chemical Engineering, Vol 2, Edition 4,p.731), a pneumatic dryer is selected to dry the crystals because the rate of drying is fast, thus protecting the product from prolonged periods of drying. The estimation calculations are shown in Appendix 8.7. The volume of the dryer is about 228 m^2, residence time is 5 sec and hot air at 125°C is used as the drying agent. Since the hot air is expensive, part of the moist hot air leaving the dryer can be recalculated and combined with the fresh air to reduce the cost. This recycle process is not included in the calculations.

8.2.7. Condenser

The vapor leaving out the evaporator is at vacuum pressure, 0.095 bar, and has to be condensed. Because of the concerns mentioned in Chapter 5.2.7, surface condenser is chosen as the type of condenser. It has exactly the same mechanism with a heat exchanger. Horizontal tube heat exchanger has been chosen for the design calculations.

Table 8.4 Design Information of Condenser

Heat Duty (kW)		24120
Process Flow (kg/hr)		35760
Temperature (°C)	In	45
	Out	37
Overall Heat Transfer Capacity (W/m*°C)		440.816
Area (m^2)		2452
Number of Tubes		21320
Tube Diameter (mm)	Inside	18
	Outside	20

The detailed information on the equipment calculations of each equipment can be found in Appendix 8.

8.2.8. Alternative Operations

The separation of the crystals is done with a press and frame filter. These are batch filters and require maintenance as well as extra operators to clean each batch of filter consuming time and elevating the cost of operation. An alternative to maintain the process continuous with less costs and a good performance is the use of a hydro-cyclone as suggested by Coulson and Richardson's to concentrate the stream and facilitate the use of a continuous filtration system. The calculation of the hydro-cyclone and the rotary drum filter that could be use for this process can be seen in Appendix 12.

For the calculation of the hydro-cyclone the nomogram 4.21 and 4.22 from page 42, Coulson & Richardson's is used, a diameter where the efficiency of the hydro-cyclone will be 50% was assumed, this diameter is set to half of the minimum diameter of the particle of crystals in order to guarantee that all the particles of the desired value are concentrated (50-150 μm).

The stream leaving the hydro-cyclone was set to 18% due to the lack of information on real hydro-cyclone processes for purposes similar to this design.
With this concentration of solids in the stream leaving the hydro-cyclone, the use of a rotary filter is a very good option to make the filtering process continuous with very good performance and less cost.

The rotary drum filter is set to operate at 0.7 bar vacuum with 30% of its area submerged. The crystals in the filter are washed and dried in the same equipment; thus, the unit operations in the downstream processing are further reduced with an effective equipment and lower capital and operational cost.

The calculation of the rotary drum filter is made using the equations. 14.2-24 found in Geankoplis, 2000.

The cost of the use of these equipment is relatively low compared to the selected equipment, the rotary drum filter is around 38906 €(Coulson & Richardson's, Vol 6) and the hydro-cyclone price is around 400 to 7000 € according to the catalog of Schumacher. Irrigation Inc. (http://www.schumacherirrigation.com, visited on 16/10/05)

The use of a hydro-cyclone for the biomass is not possible because this equipment is useful for particle sizes between 4-500 μm.

9. WASTE

In this chapter, the waste generated by the process and its disposal is discussed. The wastes generated by the processes are listed in Table 9.1, and the full inventory can be seen in Appendix 9.1.

Table 9.1 *Waste generated*

Waste	Description	Origin	Possible Effect	Disposal Method
Spent Biomass	Biomass produced in the process	<116>	Unsafe for Environment, Bad odors	Burnt in the Facilities
Spent gas	Air out of fermentor might contain biomass and (by-) products	<113>	Bad odors, Unsafe for Environment,	Through filter
Condensed Water Purge	Generated by evaporator.	<140>	Effect on ecosystem	Cooling down before discharge to the environment
Washing Water & purged filtrate	Waste from the Plate and frame filters and purged filtrate during the process	<134> <146>	Pollution of ground water	To WWTP
Spent hot gas	Air was used to drying the product	<129>	None	Release to Environment

The spent biomass, recombinant *E.coli*, should be burned because of the environmental regulations regarding modified microorganisms due to the uncertainty of its safety for the environment. This can be done internally if facilities are present[5].

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[5] The burning equipment for the biomass is considered to be out of the battery limits

The spent air should be filtered to reduce the odors and prevent the small amount of biomass spread to the surrounding environment, for the calculation purposes the possible biomass present in the spent air is not taken into consideration. Condensed water from evaporator (37°C) has high temperature and it should be cooled before it's discharged.

The washing water and purged filtrate contain certain amount of glucose, L-Phe, and acetate, thus, it is better to treat them in the WWTP before discharge. Chemical oxygen demand (COD) is used to calculate the load of aqueous waste streams and their costs. The COD of each compound is calculated using the oxidation reaction and calculating the oxygen consumption in kg O_2/kg of compound. Table 9.2 shows COD value for each component and the total wastewater COD per year. As seen in Table 9.2, COD of biomass and L-Phenylalanine take the main part of the total COD. Determining the productive biomass life cycle to possibly lengthen the production phase and as consequence reduce the batches per year with the same amount biomass per batch might reduce the biomass COD to waste. Reducing the permeate purge can decrease L-Phe amount present in the waste as it constitutes at least 80% of the total product lost.

Table 9.2 COD Load

Component	**COD kg/kg**	**COD kg/year**
Biomass	1.3659	104595
L-Phe	2.3246	318827
Glucose	1.0667	58444
Acetate	1.0667	25916
Total		**507782**

10. PROCESS SAFETY

The safety of chemical process design is very important in the industry. The general concern for safety is growing according to the development of more complex and extreme conditions of the design, the effects on the community, workers and the environment has increased the legislation and procedures for the operation of this processes. In the view of industry, there are also potential economic and business losses; therefore, process safety should be considered in every step of the design. In order to improve the safety of operation, two of the most common and convenient tools, the DOW fire and explosion index (F&EI) and a hazard and operability study (HAZOP), are used in this chapter to perform the safety analysis.

Dow F&EI is a valuable tool in the basic design stage of a project, as it is a quick method to calculate the fire and explosion intensity of a process unit, and to help in taking steps to reduce the intensity if found to be unacceptable. Moreover, it allows for estimation of the economic loss in the case of an accident. In this chapter, F&EI will be only performed on the most hazardous equipment in the process because of the time limitation.

A HAZOP is an open-ended systematic method, which can identify hazards due to fixed equipment and procedures, and provides a controlled mechanism to consider the things that can go wrong, the outcome of this method is a set of possible actions ad procedures to eliminate or avoid negative effects while improving performance, quality, and safety.

10.1. Dow F&EI Study

For the Dow F&EI study, the most hazardous equipments in the process are identified and analyzed. The pneumatic dryer and the vacuum evaporator are considered to be the most hazardous equipments in our process; therefore, the analysis was performed only on these equipments and their operations. The results of the F&EI for the pneumatic dryer and for the vacuum evaporator can be seen in Appendix 10.1 and 10.4, respectively.

According to Table 9.3 (p 369, Coulson & Richardson's), the potential degrees of hazard of the two equipments are in the light category.

10.2 HAZOP Study

A HAZOP was performed on the whole process and the results are given in Appendix 10.7. Dust explosion in the solid handling section was considered to be the one of the key safety problem. A sprinkler system would be effective for fighting fires. Preventing sparks generated in the environment or adjacent equipments, e.g., electrostatic charges can reduce explosions. The open flame in the process is an obvious source of ignition and must be sited well away from the dryer.

Overflow in the fermentor and the evaporator should be controlled according to the details listed in Appendix 10.7, in order to prevent the hazards of the recombinant biomass and water pollution by L-Phe, acetate, etc. Relief valves and instruments are needed to prevent vacuum collapse; also, the dimpled jackets may provide a way of vacuum protection when jacketed vessels are involved. Additionally, the eye showers and first aid equipments should be always present on site.

In spite of these precautions, further measures are usually required for the safety conditions of the plant. These studies and measures used in this chapter should only be seen as a preliminary evaluation, the design of a complete set of safety measures is best to left to safety experts.

In conclusion, the degree of hazard of this process is relatively low, and the process can be evaluated as a well inherent safety design.

11. ECONOMICS

In this chapter the economic aspects of a designed plant are discussed. Economic analysis has been done with the help of Coulson & Richardson Vol. 6, Chapter 6 and Bioseparations science and Engineering, Harrison R.G., Todd, P., Rudge, S.R., Petrides, D.P., 2003.

This chapter starts with the investments. In the next subchapter the operating costs, which are the costs of production are covered. The sales income of product and calculated cash flow are introduced. Then economic criteria are described and the calculated values of rate of return, pay out time and discount cash flow rate of return are given. This part of the report is important, because here the economical feasibility of the system is analyzed. A sensitivity analysis will be given taking as basis the selling price of the product as seen in Chapter 11.4 and table 11.4.

11.1. Economic Calculations

11.1.1. Investment

The total investment required for the designed plant includes the cost of fixed capital and working capital. Fixed capital is a once-only cost that is not recovered at the end of the project life. Working capital is the additional investment needed, over and above the fixed capital, to start the plant up and operate if to the point when income is earned.

To calculate the fixed capital, first the purchased equipment costs (PCE) is required to be determined. The results of the PCE are shown in Table A11.1 in Appendix 11. Then with the "Lang Method" the fixed capital can be calculated based on the known PCE.

Lang factors are selected from Bioseparations Science and Engineering Harrison R.G., 2003, where ranges of the factors are available. Working capital in this design is assumed to be 15 per cent of the fixed capital, which is the typical figure for chemical plant.

Finally, the total investment can be determined based on the known fixed capital and working capital. The calculation of the total investment is shown in Table A.11.2 in Appendix 11.1.

11.1.2 Operating Costs

The operating costs are calculated according to Coulson & Richardson's Vol 6. The cost of the production includes fixed operating costs and variable operating costs. The former do not vary with production rate while the latter do. Variable operating costs for instance include the cost of raw materials and utilities and wastes, which are shown in Table A11.3 in Appendix 11.1. The results of production costs are given in Table A11.4 in Appendix 11.1.

11.1.3. Income

Annual income is dependent on the sales volume and selling price of the product. The selling price of L-Phenylalanine is determined from DSM, which is 8.2 € per kg. The annual production rate of L-Phenylalanine is set to be 1000 tons.

11.1.4. Cash Flow

The Net Cash Flow (NCF) is calculated by subtracting the production costs from the sales income of the product. The calculation for NCF is given in Table 11.1.

11.2. Economic Criteria

In this subchapter, the relation between investment and annual income and costs is described. The project lifetime is set to be 15 years.

11.2.1 Rate of Return (ROR) and Pay Out Time (POT)

Rate of return (ROR), which is the ratio of annual profit to investment, is a simple index of the performance of the money invested. The calculation of ROR is complicated since the annual profit (NCF) will not be constant over the life of the project. A simple method is to base the ROR on the average income over the life of the project and the original investment.

$$\text{ROR}=\frac{\text{Cumulative net cash flow at end of project}}{\text{Life of project}\times\text{original investment}}\times 100 \text{ per cent} \tag{11.1}$$

The calculation of ROR and cash flow diagram is shown in Appendix A11.2. Pay out time is the time required after the start of the project to pay off the initial investment from income. It can be directly observed from Figure A11.1.

11.2.2. Discounted Cash Flow Rate of Return (DCFRR)

Discounted cash-flow analysis is used to calculate the present worth of future earnings and is sensitive to the interest rate assumed. Discounted cash-flow rate of return is an interest rate at which the cumulative net present worth at the end of the project is zero. DCFRR is a measure of the maximum rate that the project could pay and still break even by the end of the project life.

$$\sum_{n=1}^{n=t}\frac{NFW}{(1+r')^n}=0 \tag{11.2}$$

Where r'= the discounted cash-flow rate if return (per cent/100)

NFW= the future worth of the new cash flow in year n,

t = the life of the project, years

The calculation of DCFRR is shown in Table A11.6 and Figure A 11.2.

11.3 Cost Review

The final results for the economic figures are shown in Table 11.1.

Table 11. 1 Final Economic Figures

	Value **(k€)**
Purchased Equipment Costs (PCE)	**1630.5**
Investment Costs	
Direct Costs	1891
Indirect Costs	946
Direct Fixed Capital (DFC)	3121
Total Investment Costs	**3589**
Operating Costs	
Fixed Costs	755.4
Variable Costs	6749.5
Annual Production Costs	**7504.9**
Income	**8194**
Net Cash flow	**689**

Table 11.2 Comparison of Margin

	Value (k€/kg)
Margin (NCF/kg L-Phe)	0.69
Margin (From B.O.D Chapter 3.4)	5.31

The margin calculated from the Basis of Design is as 5 times as that obtained from complete analysis since the latter includes fixed operating costs and other operating costs (see Table 11.1), which is not included in the B.O.D. design.

With the economic figures the economic criteria can be determined and shown in the Table 11.3.

Table 11.3 Economic Criteria

	Value
Life of project (k€)	15
Net cash flow (k€/a)	689
Original investment (k€)	3586.3
ROR (%)	19.2
POT (years)	9
DCFRR (%)	12.6

This project is economically feasible based on the complete economic analysis:
With the selling price of 8.2€ /kg of L-phenylalanine, economic margin of 0.69€ /kg of product, rate of return of 19.2% and discounted cash flow rate of return of 12.6% can be obtained. Based on this selling price, it will take 9 years for the project to pay back the investment. The project will earn a return on the investment since 9 years after the beginning of the life of the project.

11.4. Sensitivities

A sensitivity analysis is to examine the effects of uncertainties in the forecasts on the viability of the project. In this project, an error of ±10% on the sales price is assumed. This will show how sensitive the cash flows and economic criteria are to the errors in the forecast figures.

Table 11.4 Sensitivity Analysis Results

	Case 1 (7.4€/kg)	**Case 2** (8.2€/kg)	**Case 3** (9.0€/kg)
Net cash flow (k€/a)	-130	689	1509
ROR (%)	-	19.2	42.1
POT (years)	>15	9	5
DCFRR (%)	-	12.6	25.8

The selling cost of L-Phenylalanine is set to be 10% lower than 8.2€ per kg in case 1. In case 2 the selling cost of the product is 8.2€ per kg. Then in case 3, the selling cost is set to be 10% higher than 8.2€ per kg. The comparison of cumulative cash flow among these three cases is shown in Figure 11.1.

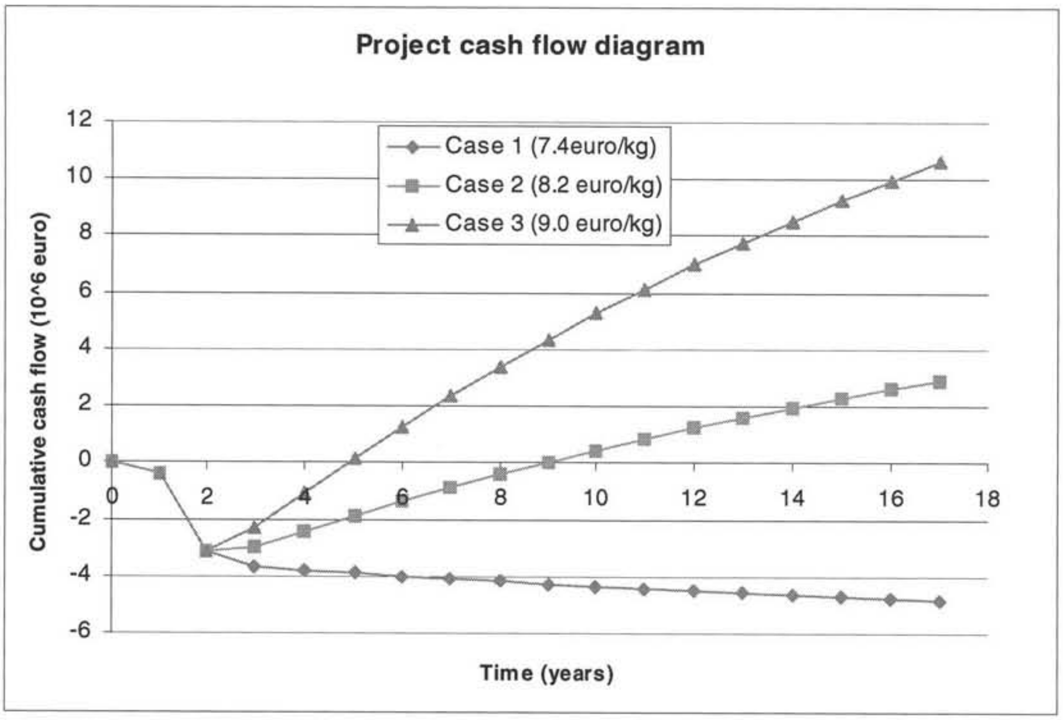

Figure11.1 Cumulative cash-flow diagram for the comparison of three cases

It is obtained from Table 11.4 and Figure 11.1 that an increase of the selling price of L-Phenylalanine by 10 per cent will increase the ROR from 19.2% to 42.1%, DCFRR from 12.6% to 25.8%, while decrease the pay out time from 9 years to 5 years. On the contrary, a decrease of the selling price of L-Phenylalanine by 10 per cent will result in negative cash flow and the investment cannot be paid back until the end of the life of the project.

12. CREATIVITY AND GROUP PROCESS TOOLS

In the creativity studies Prof.Henk Nugteren has helped to the group in conceptual process design project. In this part the studies made with him will be explained and positive contribution to the efficiency in-group work will be described. Because of being an international design group with members from different countries and cultures the creativity meetings were always very interesting and useful to understand each other with the guidance and expertise of our creativity coach Prof. Henk Nugteren.

The first meeting with the creativity coach, Prof.Henk Nugteren, has been taken place on August 9. In the meeting the essays are submitted and in light of the read articles the applicable process tools to the group work are discussed which are: Free association, false relation, brainstorming, brain-writing, sketching, and thinking sketch.

In the next appointment in the other week, the impressions obtained from the articles are discussed and the discussion focuses on the tools found useful by the members of the group. Prof. Nugteren has also helped the group how these methodologies should be realized during group work.

The other week a brain storming session has been done in the presence of the creativity coach to find solutions on the crystallization of other components in the system and to minimize their negative effect to L-Phenylalanine crystallization.

In one of the meetings, the alternatives to increase the evaporation efficiency have been discussed, due to requirement for high amount of evaporation. Two solutions are found to the problem: use of an external evaporator and use of spray crystallizer.

As a group we thanks to our creativity coach for sharing his time with us.

13. CONCLUSION AND RECOMMENDATIONS

- Determining the productive biomass life cycle to possibly lengthen the production phase and as consequence reduce the batches per year with the same amount biomass per batch. This might reduce the biomass COD to waste, reduce the costs and improve production.

- Determine the effect that different compounds in the fermentor might affect the crystallization of the product in order to optimize the operation of the system.

- The use of a second crystallization step to recover the dissolved L-Phenylalanine wastes in the purge, this amount represents around 80% of the total product lost.

- ·Study the use of immobilized biomass in the fermentor to reduce the separation processes and lengthen the time span of each batch.

- In the crystallizer acetate, glucose and ammonia are present and its effect on the solubility of the L-Phenylalanine is not known. This may cause an increase in its solubility leading product cost and higher recycle load.

- Heat integration seems to be obligatory to the system to decrease the cost of utilities. Especially condenser (E102) has a high utility consumption with around $1x10^6$ kg/hr where 35760 kg/hr steam with high potential of energy is tried to be condensed. At the first step in heat integration this amount of steam is considered to be used in another place of the process to heat process streams. However the low temperature of the steam, 45oC, does not allow this operation where process liquids will be heated until that temperature.

- This project is economically feasible based on the complete economic analysis: With the selling price of 8.2€ /kg of L-phenylalanine, economic margin of 0.69€ /kg of product, rate of return of 19.2% and discounted cash flow rate of return of 12.6% can be obtained. Based on this selling price, it will take 9 years for the project to pay back the investment. The project will earn a return on the investment since 9 years after the beginning of the life of the project.

- Selling income and annually production cost are the two main factors that have impact on the new cash flow, the economic margin of the product and also other economic criteria. Sensitivity analysis based on the fluctuation of the selling price by 10% shows that economic criteria are sensitive to the selling price due to huge quantity of the annual production rate in this project. Reducing the desired price by 10% will result in negative cash flow and the investment cannot be paid back until the end of the life of the project. Hence, fluctuation on the selling price will bring a certain degree of risk to the forecast performance of the project.

- As seen in the calculations the main factor affecting the production cost are the utilities. This is due to the characteristics of the equipments that require high amounts of heat. The highest amount of utility consumption is observed in condenser to condense the steam evaporated, approximately 1×10^6 kg/hr cooling water is required. Instead of trying to cool down the steam at low temperatures, 45°C, the design could be changed and the steam produced can be accounted as a loss while replacing that amount of water through evaporation by feeding the same amount water to the fermentor.

- The economic margin of the product is 0.69€ /kg when the selling price of the L-Phenylalanine is set to be 8.2€ /kg. As stated by Bongaerts *et al.*, 2001 the selling price for L-Phenyalanine is 20-24 US$/kg (16.4-19.66€ /kg) almost double the selling price used in the present work. If a higher margin of profit is desired the selling price could be increased according to the criterion of the client to be competitive and get higher profits.

- On the other hands, the chosen price is low to make the company a good entrance to the market. Taking account the success in the market in the first years, the price of the product can be increased to obtain higher profits. In this way the company gets a higher strength to compete in market

LIST OF SYMBOLS

Symbol Description		SI Units
A_{cross}	Cross-sectional Area	m^2
A_m	Membrane area	m^2
$A_{perpass}$	Cross-sectional Area per Pass	m^2
cv	Heat of vaporization	J/kg/°C
C	Concentration	gm/L
Cp	Heat capacity	J/mol/K
d	Diameter	m
D	Diameter vessel	m
D_b	Bundle Diameter	m
D_c	Bundle Diameter Clearance	m
D.s	Stirrer diameter	m
d_e	Equivalent Diameter	m
d_w	Wall thickness cooling coil [m]	
D_i, D_o	Inside and Outside Diameters	m
D_s	Shell Inside Diameter	m
f	Mass fraction	-
f_c	Convection Correction Coefficient	-
F	Stream Mass Vector	kg/hr
F_{IN}	Stream Mass Flow of Enterance Stream	kg/hr
$Ftotal$	Stream mass flow	kg/hr
g	Gravity	m/s2
G	Gibbs energy	J/mol/K
G	Growth rate of crystals	m/s
h	Disk spacing	m
hc_BK	Convective Heat Tranfer Coefficient Corrected by Boyko-Kruzhilin Correlation	W/m^2*°C

hi	Inside Convective Heat Transfer Coefficient	W/m^2*^oC
hs	Shellside Convective Heat Transfer Coefficient	W/m^2*^oC
h_l	Heat transfer coefficient medium	$W/m^2/K$
H	Enthalpy	J/mol/K
H	Vessel Height	m
Had	Adiabatic Head	J/kg
j_f	Friction factor	-
J	Permeate flux	$m^3/m^2/hr$
k	Thermal Conductivity	$W/m*^oC$
kla	Oxygen transfer coefficient	1/hr
K	Cp/Cv	
Ki	Inhibition constant	gm/L
Ks	Affiniti constant	gm/L
Kg	Growth rate constant	m/s
L	Tube Length	m
L_suction	Pipe length in suction side [m]	m
l_B	Baffle Spacing	m
Lm	Mean size of crystals	m
mf	Mass fractions centrifuge	-
M	Mass	kg
Mw	Molecula weight	gm/mol
m_o	Herry coefficient of oxygen	-
Nt	Number of Tubes	-
OUR	Oxygen uptake rate	$molO_2/m^3/hr$
P	Pressure	bar, atm
Pf	Pressure loss in the suction piping [N/m^2]	[N/m2]
P_ov	Power input by stirring,	W/m^3
Pv	Vapour pressure of the fluid [N/m2]	[N/m2]
q	Production/Consumption	g/gX/hr
q	Heat Flux	W/m^2*^oC

qm	Manteinande coefficeint	g/gX/hr
Q	Heat Duty	kJ/hr
Qevap	Evaporator heat	W
Qc	Crystallizer heat	W
r	Rate of production/consumption	gm/hr
r	radius	m
rh	Heat	W
R	Ideal gas constant	J/mol/K
Re	Reynolds Numer	-
R_m	membrane resistance	1/m
S	Solubility	gm/L
S	Supersaturation	-
t_f	Filtration time	hr
T	Temperature	K
T_1	Inlet Shell Side Fluid Temperature	^{o}C
T_2	Outlet Shell Side Fluid Temperature	^{o}C
t_1	Inlet Tube Side Fluid Temperature	^{o}C
t_2	Outlet Tube Side Fluid Temperature	^{o}C
T_S	Steam Temperature	^{o}C
U	Overall Heat Transfer Coefficient	W/m^2*^{o}C
u	Velocity	m/s
u_t	Tube-side Velocity	m/s
V	Volume	m3
Visc	Viscosity	Pa*s
v_{vol}	Volumetric Flow	m^3/s
vs	Superficial gas velocity	m/s
w	Recycle fractions	-
W	Evaporated Water Mass Flow	kg/hr
Y	Yield	-
x	Vapor Fraction	-

xo	Oxygen fraction in gas	-
Xtt	Lockhart-Martinelli Two-phase Flow Parameter	-
Z	Number of disks	-
Greek	Description	SI unit
DP_S	Shell-side Pressure Difference	bar
DT_{Lm}	Logarithmic Mean Temperature	^oC
G	Tube Loading	kg/m*s
a_c	Surface area cooling coils	m^2
a	Specific cake resistance	m/kg
l	Latent Heat	J/kg
l_w	Heat conductivity of the well	W/m/K
l_b	Heat conductivity of the broth	W/(m*K)
m	Growth rate	1/hr
r	Density	kg/m3
D	Change	
t	Residence time	s
f	Flow	kg/h, m3/h
e	Specific power dissipation	W/kg
n_b	Kinematic viscosity of the broth	m^2/s
h_b	Dynamic viscosity of the broth.	Ns/m^2
η	Pump efficiency	[%]
e/d	Relative roughness	-
Δp_f	Pressure drop in a pipe due to friction	[N/m2]
Δz	Difference in elevations	[m]
ΔP	Difference in system pressures	[N/m2]

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APPENDIX 1 PRE-PRODUCTION PHASE

With the definition of pre-production the required phases to reach the steady state production phase are described. Related calculations with this part can be found in CD under file (FermentorPhaseCalculations.mcd). Only the important related results will be shown here.

Appendix 1.1. Inoculum

For the pre-cultivation step we will assume the method used by Maass et al, 2002. Also as described in the same article, the precultivation media is composed to be on the glucose and tyrosine concentration. The volume to be inoculated is set to be 10% of the working volume of the fermentor, which is equal to 8 m^3. Total volume of the precultivation tank is found to be 16 m^3. The height over diameter ratio is set to be 2. The size of the inoculum tank is calculated as the following: Diameter: 1.366 m ; Height: 2.731 m. The MathCad file including the calculations can be found in CD under file incoculum.mcd.

Appendix 1.2. Batch Phase

- Biomass Concentration

$$\text{Cx\_batch}(t) := \text{Cx0}\exp\left(t\cdot\mu_{\text{max}}\right)$$

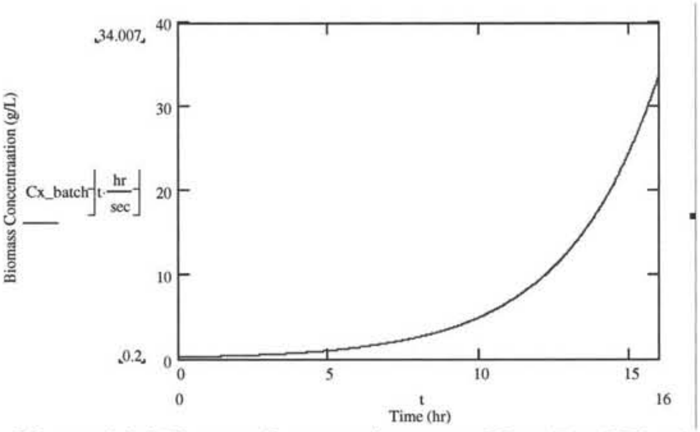

Figure A.1.1 Biomass Concentration versus Time (Batch Phase)

- Phenylalanine Concentration

$$\mathrm{Cphe\_batch}(t) := \mathrm{Cp0} + \mathrm{q\_Phe}\cdot \mathrm{Cx0}\cdot\left(\frac{\exp(\mu\cdot t)}{\mu_{\max}} - \frac{1}{\mu_{\max}}\right)$$

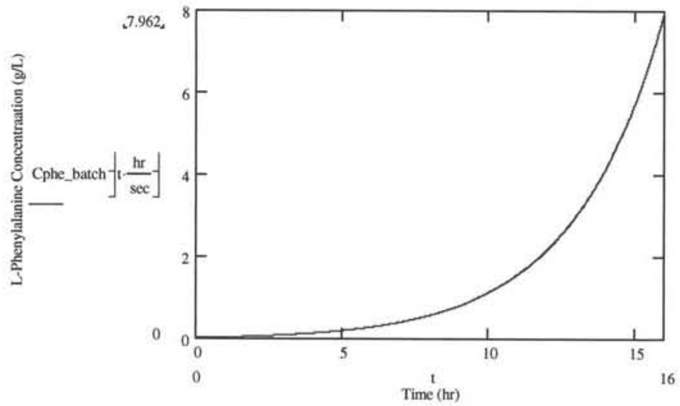

Figure A.1.2 Phenylalanine Concentration versus Time (Batch Phase)

- Tyrosine Concentration

$$\mathrm{Ctyr\_batch}(t) := \mathrm{Ctyr0} + \mathrm{q\_Tyr}\cdot \mathrm{Cx0}\cdot\left(\frac{\exp(\mu\cdot t)}{\mu_{\max}} - \frac{1}{\mu_{\max}}\right)$$

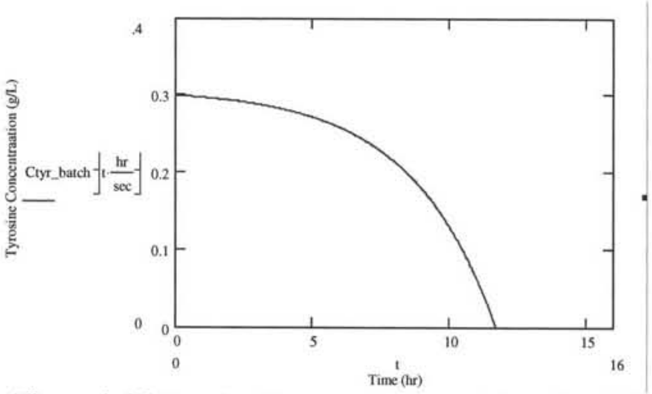

Figure A.1.3 Tyrosine Concentration versus Time (Batch Phase)

- Glucose Concentration

$$\mathrm{Cglu\_batch}(t) := \mathrm{Cglu0} + \mathrm{q\_Glu}\cdot \mathrm{Cx0}\cdot\left(\frac{\exp(\mu\cdot t)}{\mu_{\max}} - \frac{1}{\mu_{\max}}\right)$$

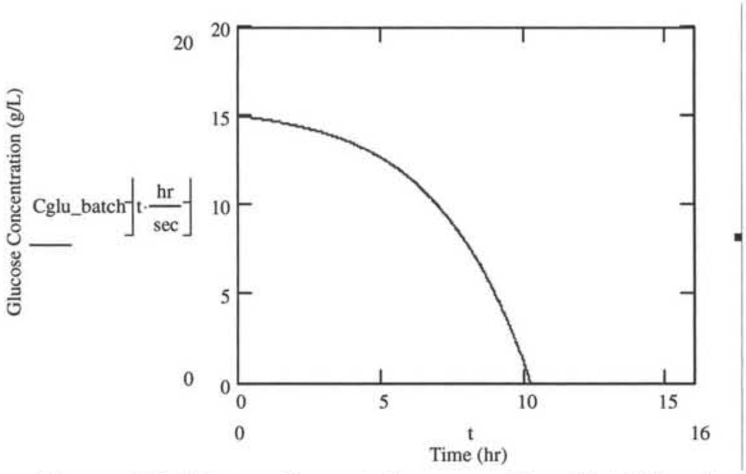

Figure A.1.4 Glucose Concentration versus Time (Batch Phase)

Glucose concentration drops to 5 g/L 8.976 hours later (8hrs56min). Glucose feed starts to be fed. In other words fed-batch extended growth phase starts

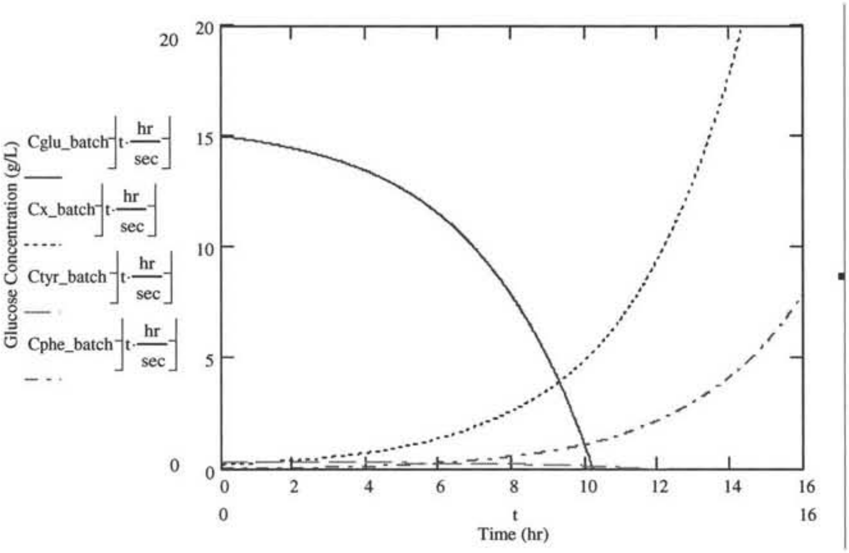

Figure A.1.5 Concentrations of all components versus Time (Batch Phase)

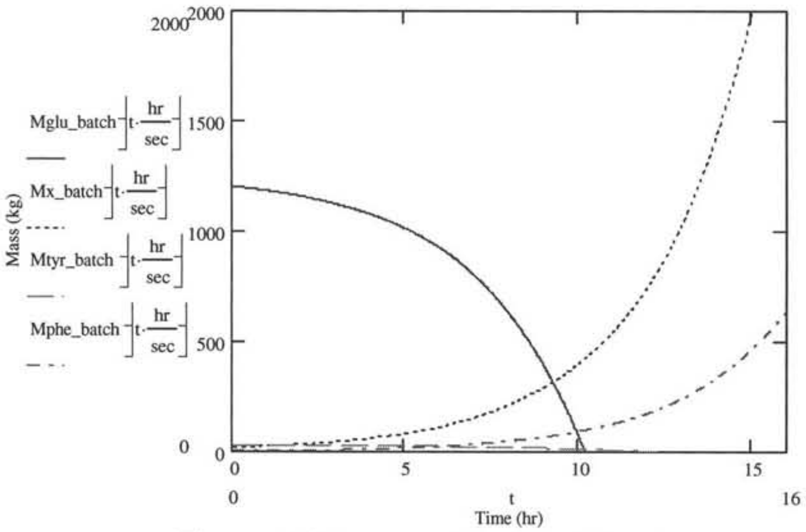

Figure A.1.6 Mass versus Time (Batch Phase)

Conditons at the End of Batch Phase:

$V = 80 m^3$

$Cx\_batch(T_{Batch}) = 3.567$ g/L

$Cglu\_batch(T_{Batch}) = 4.982$ g/L

$Ctyr\_batch(T_{Batch}) = 0.179$ g/L

$Cphe\_batch(T_{Batch}) = 0.793$ g/L

$Mx\_batch(T_{Batch}) = 285.398 kg$

$Mglu\_batch(T_{Batch}) = 398.583 kg$

$Mtyr\_batch(T_{Batch}) = 14.31 kg$

$Mphe\_batch(T_{Batch}) = 63.447 kg$

Raw Material Consumption at Batch Phase:

- Glucose : 801.417 kg
- Tyrosine : 9.69 kg
- Ammonium : 2.618 kg

Appendix 1.3. Extended Growth Phase

As can be seen in the graph representing the concentration changes in batch phase, glucose concentration reaches to the set value 5 g/L after 8.976 hours batch starts, whereas Tyrosine requires more time to reach the set value 0.01 g/L.

Therefore first Glucose must be fed first to the system. After tyrosine level reaches 0.01 g/L also tyrosine feed will be fed. From this point of view extended growth phase can be divided into two parts.

Part - 1

- Determination of time of extended growth phase part 1

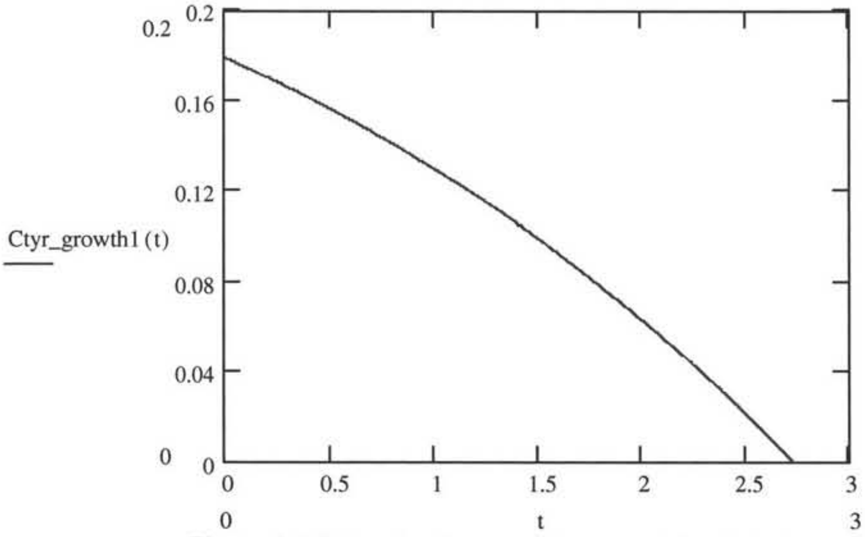

Figure A.1.7 Tyrosine Concentration versus Time (Ext. Growth-I)

After 2.61 hours of extended growth phase operation Tyrosine concentration falls down to 0.01 g/L. At that time tyrosine starts to be fed to the system

- Concentrations of all material

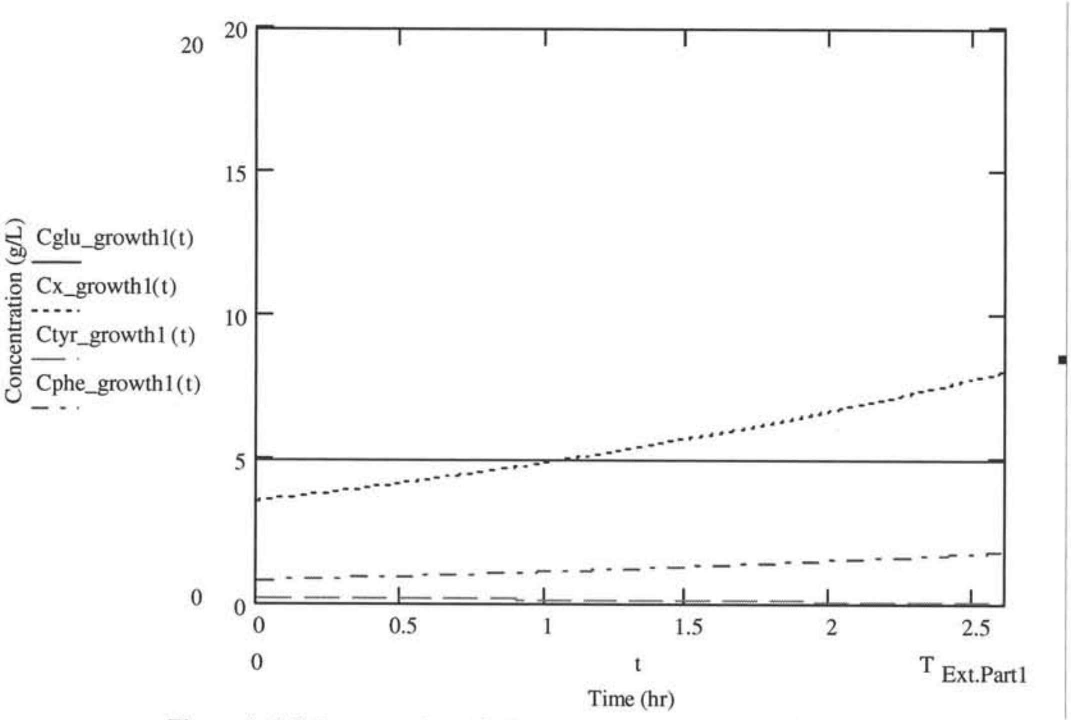

Figure A.1.8 Concentration of all components versus Time (Ext. Growth-I)

- Mass of Components

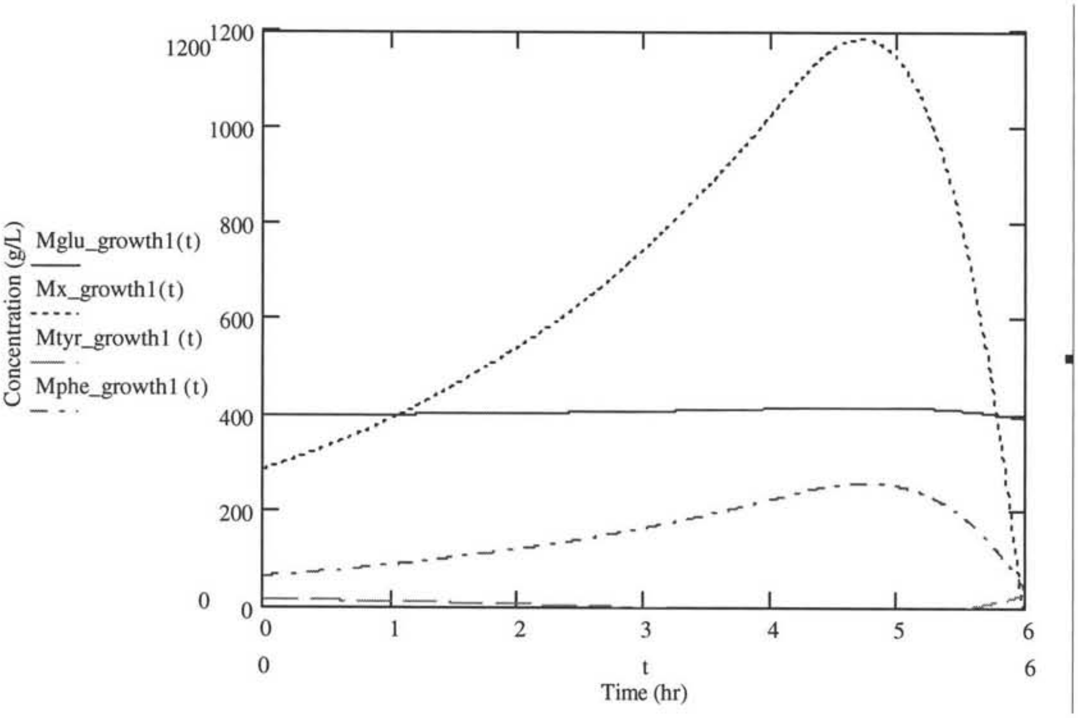

Figure A.1.9 Mass versus Time (Ext. Growth-I)

- Glucose Feed

$\Phi_{glu1}(t) := -qglu\_growth1(t) \cdot Mx\_growth1(t) \cdot kg$

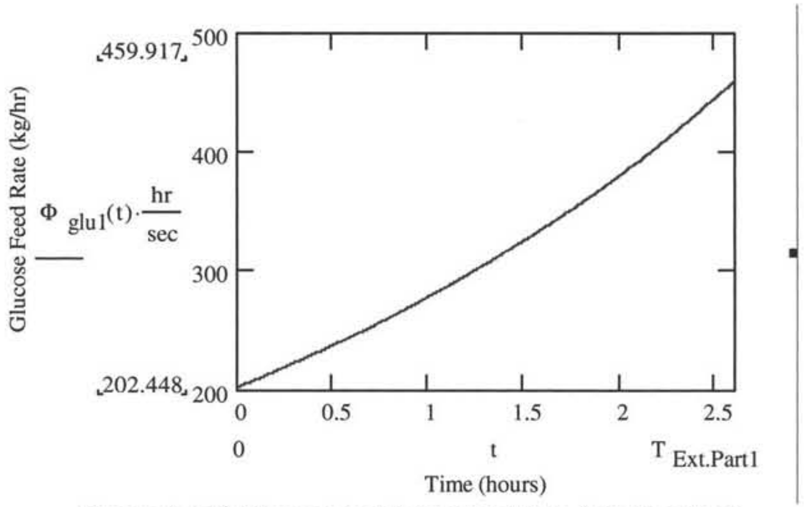

Figure A.1.10 Glucose Feed Rate versus Time (Ext. Growth-I)

- If Tyrosine is not supplied biomass would die after 6 hours.

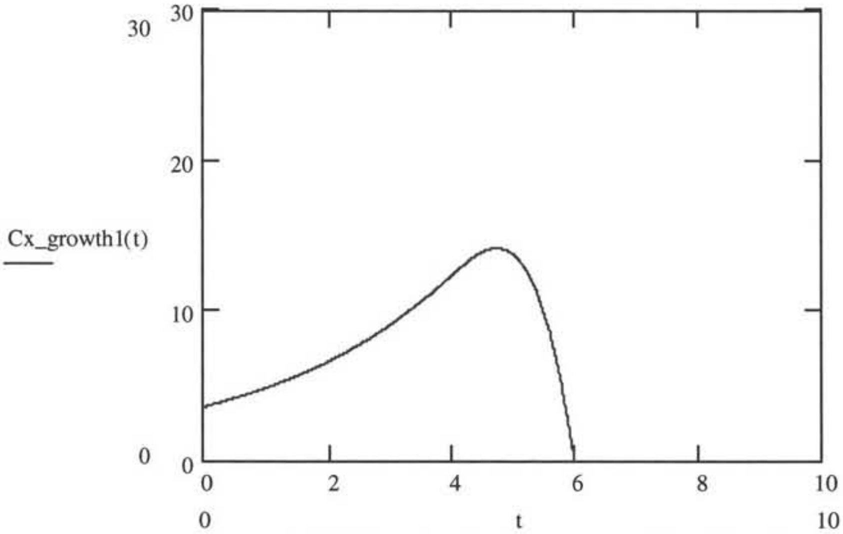

Figure A.1.11 Tyrosine Feed Rate versus Time (Ext. Growth-I)

Conditons at the End of Extended Growth Part 1:

$V\_growth1(T_{Ext.Part1}) = 81.563$	m^3
$Cx\_growth1(T_{Ext.Part1}) = 8.088$	g/L
$Cglu\_growth1(T_{Ext.Part1}) = 4.982$	g/L
$Ctyr\_growth1(T_{Ext.Part1}) = 0.01$	g/L
$Cphe\_growth1(T_{Ext.Part1}) = 1.793$	g/L
$Mx\_growth1(T_{Ext.Part1}) = 659.64$	kg
$Mglu\_growth1(T_{Ext.Part1}) = 406.368$	kg
$Mtyr\_growth1(T_{Ext.Part1}) = 0.849$	kg
$Mphe\_growth1(T_{Ext.Part1}) = 146.222$	kg

Raw Material Consumption at Extended Growth Part 1:

- Glucose : 1094 kg
- Tyrosine : 13.461 kg
- Ammonium: 3.052 kg

Part - 2

- Determination of time of extended growth phase part 2:

Extended Growth phase will be ended when 30 g/L Biomass Concentration is reached.

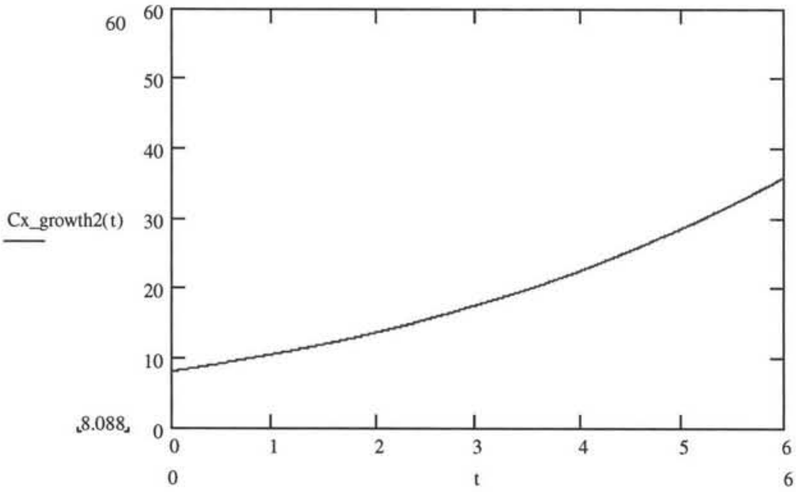

Figure A.1.12 Biomass versus Time (Ext. Growth-II)

After 5.232 hours, Extended Growth phase part 2 operation Biomass Concentration reaches to the desired value, and production phase starts.

- Concentrations of all components

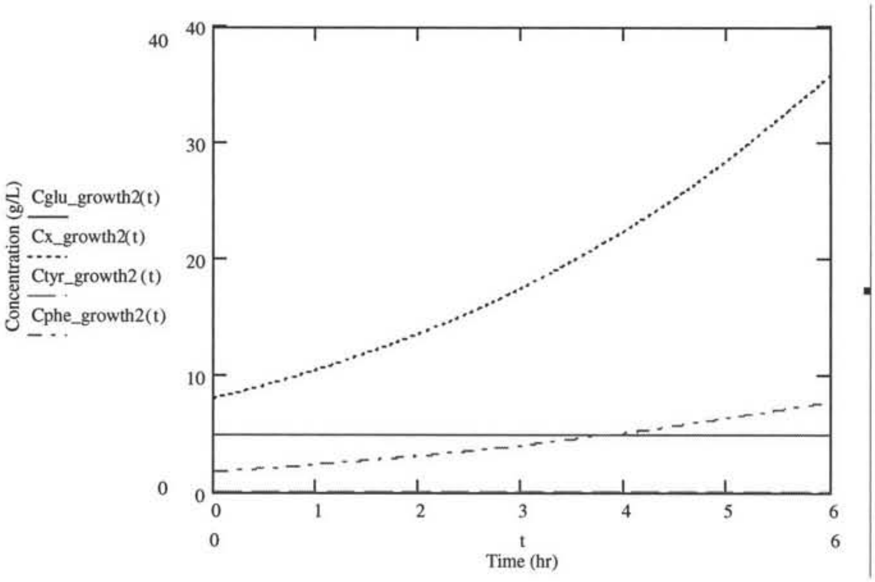

Figure A.1.13 Concentration of all components versus Time (Ext. Growth-II)

- Mass of all components.

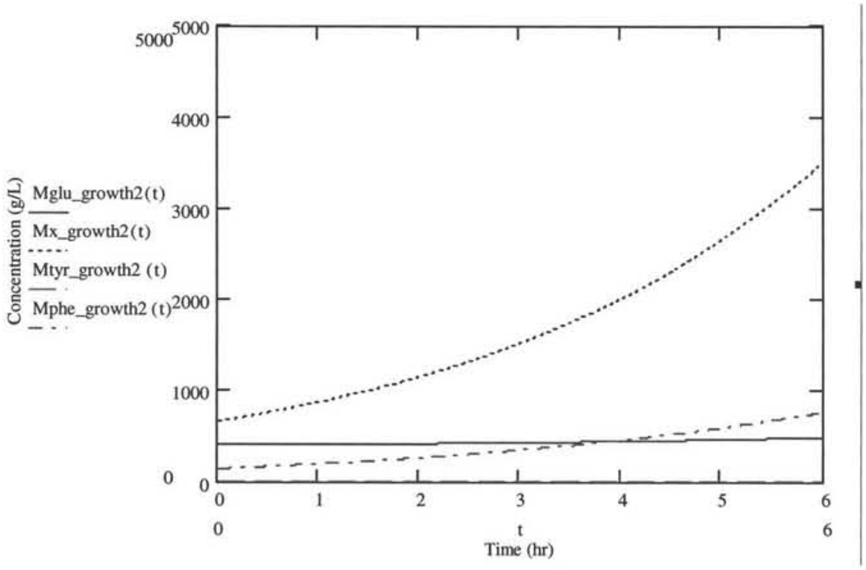

Figure A.1.14 Mass versus Time (Ext. Growth-II)

- Glucose feed

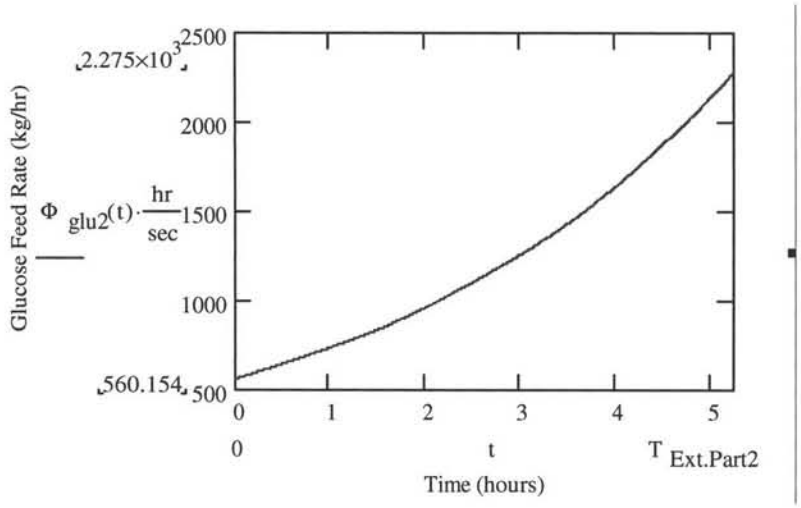

Figure A.1.15 Glucose Feed Rate versus Time (Ext. Growth-II)

- Tyrosine feed

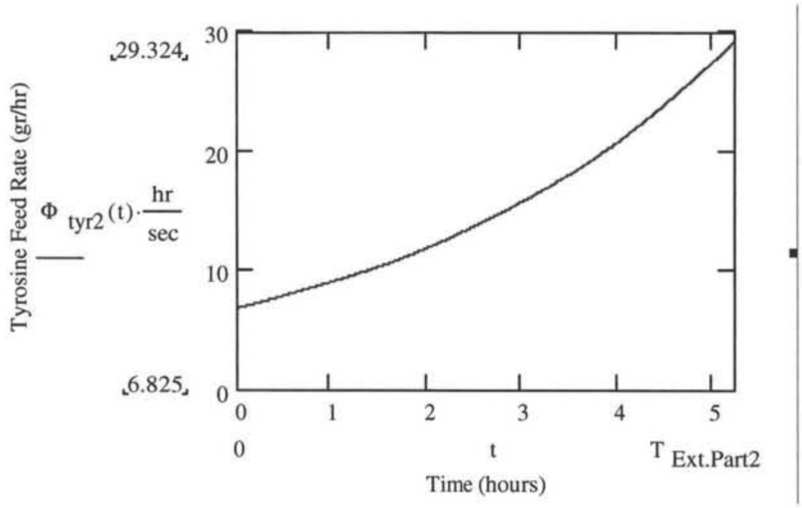

Figure A.1.16 Tyrosine Feed Rate versus Time (Ext. Growth-II)

Conditons at the End of Extended Growth Part 2:

$V\_growth2(T_{Ext.Part2}) = 93.962$ m^3

$Cx\_growth2(T_{Ext.Part2}) = 30.165$ g/L

$Cglu\_growth2(T_{Ext.Part2}) = 4.982$ g/L

$Ctyr\_growth2(T_{Ext.Part2}) = 0.01$ g/L

$Cphe\_growth2(T_{Ext.Part2}) = 6.709$ g/L

$Mx\_growth2(T_{Ext.Part2}) = 2.834 \times 10^3$ kg

$Mglu\_growth2(T_{Ext.Part2}) = 468.147$ kg

$Mtyr\_growth2(T_{Ext.Part2}) = 0.978$ kg

$Mphe\_growth2(T_{Ext.Part2}) = 630.351$ kg

Raw Material Consumption at Extended Growth Part 2:

- Glucose : 6419 kg
- Tyrosine : 80.745 kg
- Ammonium : 18.075 kg

Appendix 1.4. Production Phase

In production phase, glucose and tyrosine concentrations are kept constant at 5 g/L and 0.01 g/L during the process, respectively. The values of the feed streams are given in the mass balances and the concentration constraints are described (Chapter 7 Mass and Heat Balances).

APPENDIX 2. PROCESS OPTIONS AND SELECTION

For the decision of the best process to develop several options were analyzed and evaluated according in order to decide the one that will offer the best alternative to achieve the goals set in the description of the process. The processes evaluated are the following:

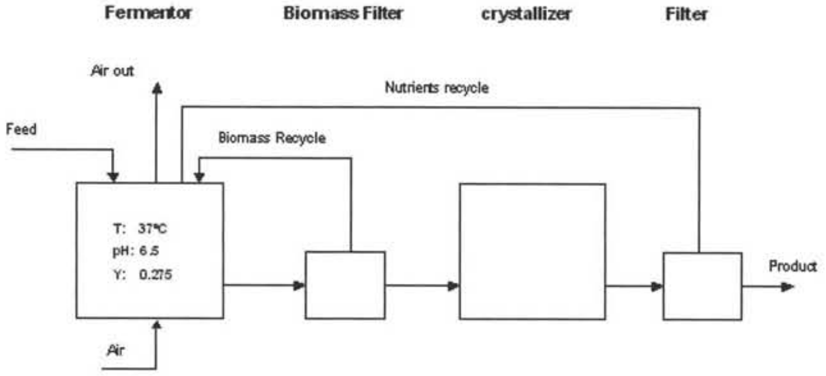

Figure A2.1 In-Situ Crystallization-Fermentation

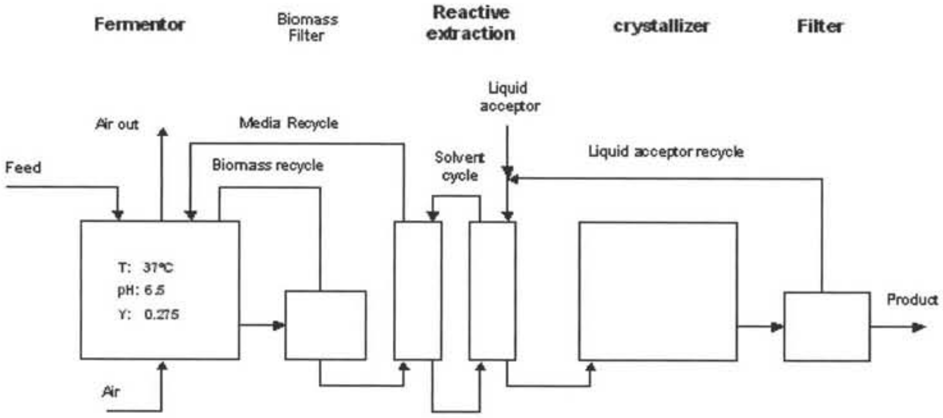

Figure A2.2 In-Situ Reactive Extraction- Fermentation Process

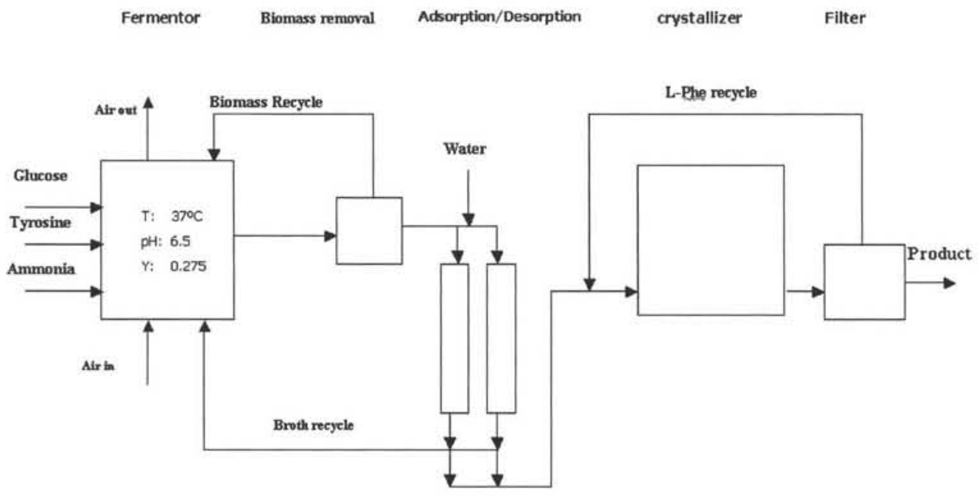

Figure A2.3 In-Situ Adsorption-Fermentation Process with the Use of Beads

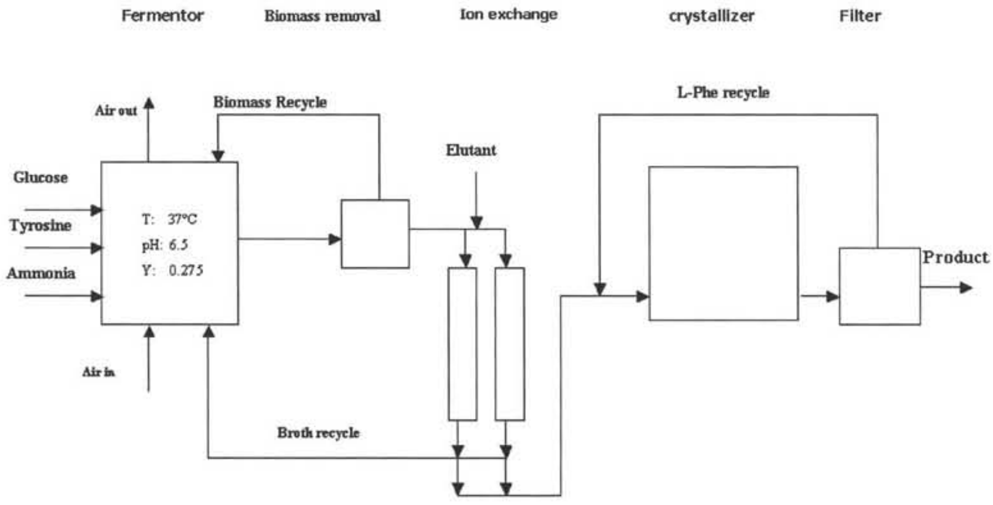

Figure A2.4 In-Situ Ion Exchange-Fermentation Process

The selected process use the same principle of ISPR of the target product, thus reducing the concentration in the fermentor. The last two process are similar in their scheme but differ in the adsorbent an separation principle for the process. Figure A2.3 uses polymeric beads(non polar) for the adsorption of the target compound while Figure A2.4 uses an ion exchange resin.

The selected process to develop is the integrated fermentation-crystallization ISPR in Figure A2.1 after the evaluation done taking in consideration the criterions stated in Chapter 2.

APPENDIX.3. BASIS OF DESIGN (BOD)

Appendix 3.1 List of pure component properties

Table A.3.1 Pure Component Properties

Component Name		**Technological Data**					**Medical Data**		**NOTE**
Design	**Systematic**	**Formula**	**MW g/mol**	**Bp °C (1)**	**Mp °C (1)**	**Density kg/m3**	**MAC mg/m3**	**LD50 g**	
L-phe	L-phe	C9H11O2N	165.19	376.45	172.82	n.a	n.a	220	(2)
Tyrosine	Tyrosine	$C_9H_{11}NO_3$	181.1908	456.85	283.85	n.a	n.a	n.a	
Glucose	Glucose	C6H12O6	180.16	571.29	169.83	1540	n.a	25.8	
Acetic acid	Acetic acid	$C_2H_4O_2$	60.0524	118.1	16.5	1049	n.a	4.96	(3)
Water	Water	H2O	18	0	100	1000	n.a	n.a	
Ammonia	Ammonia	NH3	17	-33	-78	682	n.a	0.35	(4)
Oxygen	Oxygen	O2	32	-182.96	-218.5	1.429	n.a	100 pph/14H	(5), (6)
Nitrogen	Nitrogen	N2	28	n.a	-210	1.2506	n.a	n.a	
Carbon dioxide	Carbon dioxide	CO2	44	n.a	-78.5	1.53	n.a	10 pph/min	(6)
Notes	(1) at 1 bar (2) TDLo oral rat (3) (a) TA (4) (D) at BP (5) (D) at gas (6) (TCLo) inhale								

Appendix 3.2 Battery Limit

Table A3.2 Ingoing E.coli

Name:	Ingoing E.coli				
Comp.	Units	Specification		Notes	Additional Information (also ref. note numbers)
		Available	Design		
E.coli	%wt		N/A	(1)	(1) E.Coli is modified, tyrosine-autotrophic strain. The composition of this modified strain will be provided by Principal.
Water			N/A		
Total			N/A		
Process Conditions and Price					
Temp.	°C	ambient			
Press.	Bara	1			
Phase	V/L/S	S			
Price	USD/ton	N/A		(2)	(2) Neglected, the preinoculum will be prepared on site.

Table A3.3 Ingoing glucose solution

Name:	Ingoing glucose				
Comp.	Units	Specification		Notes	Additional Information (also ref. note numbers)
		Available	Design		
Glucose	%wt	54.2	54.2	(1)	(1) Glucose is added into the fermentor as the carbon source for the biomass.
Water	%wt	45.8	45.8		
Total		100	100.0		
Process Conditions and Price					(2) The price of **pure** glucose is given by market price[1]
Temp.	°C	ambient			
Press.	Bara	1			
Phase	V/L/S	L			
Price	€/ton	245		(2)	

1 http://www.starch.dk/isi/market/company.htm#Sellers, visited 03-08-2005

Table A3.4 Ingoing tyrosine solution

Name:	Ingoing tyrosine				
Comp.	Units	Specification		Notes	Additional Information (also ref. note numbers)
		Available	Design		
Tyrosine	%wt	2.44	2.44	(1)	(1) Tyrosine is added into the fermentor as the limiting substrate for the biomass in the production phase.
Impurities		97.56	97.56		
Total		100	100.0		
Process Conditions and Price					
Temp.	°C	ambient			
Press.	Bara	1			(2) The price is given by market price of tyrosine[2] and water.
Phase	V/L/S	L			
Price	€/ton	4097.26		(2)	

2 Johannes Bongaerts, Metabolic Engineering for Microbial Production of Aromatic Amino Acids and Derived Compounds, 2001

Table A3.5 Ingoing ammonia solution

Name:	Ingoing ammonia				
Comp.	Units	Specification		Notes	Additional Information (also ref. note numbers)
		Available	Design		
Ammonia	%wt	25	25	(1)	(1)Ammonia water is applied for pH control in the fermentor.
Water	%wt	75	75		
Total		100	100.0		
Process Conditions and Price					
Temp.	°C	ambient			
Press.	Bara	1			
Phase	V/L/S	L			
Price	€/ton	0.0563			

3. http://www.chemicalmarketreporter.com

Table A3.6 Ingoing filtered air

Name:	Ingoing filtered air				
Comp.	Units	Specification		Notes	Additional Information (also ref. note numbers)
		Available	Design		
Nitrogen	%vol	52.5	52.5		(1)Price indication is obtained from[3]
Oxygen		47.4	47.4		
Other		0.1	0.1		
Total		100	100.0		
Process Conditions and Price					
Temp.	°C	ambient			
Press.	Bara	1			
Phase	V/L/S	V			
Price	USD/m^3	0.005		(1)	

4. R.G. Harrison, P. Todd, S.R. Rudge, D.P. Petrides, Bioseparations Science and Engineering (Oxford University Press,
New York, 2003).

Table A3.7 Ingoing mineralized water

Name:	Ingoing mineralized water				
Comp.	Units	Specification		Notes	Additional Information (also ref. note numbers)
		Available	Design		
Water	%wt	100	100		(1)Price indication is obtained from[4]
Total			100.0		
Process Conditions and Price					
Temp.	°C	ambient			
Press.	Bara	1			
Phase	V/L/S	L			
Price	€/ton	0.015		(1)	

5. R.K. Sinnott, Coulson & Richardson's Chemical Engineering

Table A3.8 Outgoing L-Phenylalanine

Name:	Outgoing L-Phenylalanine				
Comp.	Units	Specification			Additional Information
		Available	Design	Notes	(also ref. note numbers)
L-Phe	%wt		98.5		(1) Price set by the client
Impurities			1.5		
Total			100.0		
Process Conditions and Price					
Temp.	°C		ambient		
Press.	Bara		1		
Phase	V/L/S		S		
Price	€/ton		8200	(1)	

Table A3.9 Outgoing spent air

Name:	Outgoing spent air				
Comp.	Units	Specification			Additional Information
		Available	Design	Notes	(also ref. note numbers)
Nitrogen	%vol	N/A	78.1		
Oxygen		N/A	<21.0		
CO_2		N/A	>0		
Water		N/A	<<1		
Glucose		N/A	<<0.1		
Tyrosine		N/A	<<0.1		
Acetate		N/A	<<0.1		
L-Phe		N/A	<<0.1		
Total			100.0		
Process Conditions and Price					
Temp.	°C		<125		
Press.	Bara		1		
Phase	V/L/S		V		
Price	USD/m^3		N/A		

Table A3.10 Outgoing to WWTP

Name:	Outgoing spent process water				
Comp.	Units	Specification			Additional Information
		Available	Design	Notes	(also ref. note numbers)
Water	%wt		N/A	(1)	(1) Spent process water may contain glucose, tyrosine, acetate, fermentation chemicals. The specific information will be provided after complete calculation.
Contaminations			N/A		
Total			100.0		
Process Conditions and Price					
Temp.	°C		25		
Press.	Bara		1		
Phase	V/L/S		L		
Price	USD/ton		300	(2)	(2) Price indication is from[6]

7.Price of ton COD obtained from Prof. Dr.ir. M.van Loosdrecht, TU Delft

Table A3.11 Outgoing spent E.coli

Name:	Outgoing spent E.coli				
Comp.	Units	Specification			Additional Information
		Available	Design	Notes	(also ref. note numbers)
E.coli	%wt		N/A	(1)	(1) There is no information available for the composition of the spent biomass at this moment. Further specifications for discarding biomass should be done
Water			N/A		
Total			100.0		
Process Conditions and Price					
Temp.	°C		37		
Press.	Bara		1		
Phase	V/L/S		S		
Price	USD/ton		N/A	(2)	(2) Not yet known. Will be provided by Principal later

Appendix 3.3 Improvements in the Selected Process

Here the selected process with the improvements is shown. The new process introduces two recycle streams, one is the vapor coming from the evaporator to the fermentor, the second is a recycle stream from the permeate from the L-phenylalanine filter going back to the evaporator.

These changes were considered because the quantity of evaporated water was considerable and the waste would have been very high. The Permeate recycle was introduces in order to ke advantage of the component present in the streams such as a high concentration on L-Phe and the remaining nutrients.

Two purges were added to avoid the accumulation of unwanted compounds in the process, such as acetate, that might affect its performance

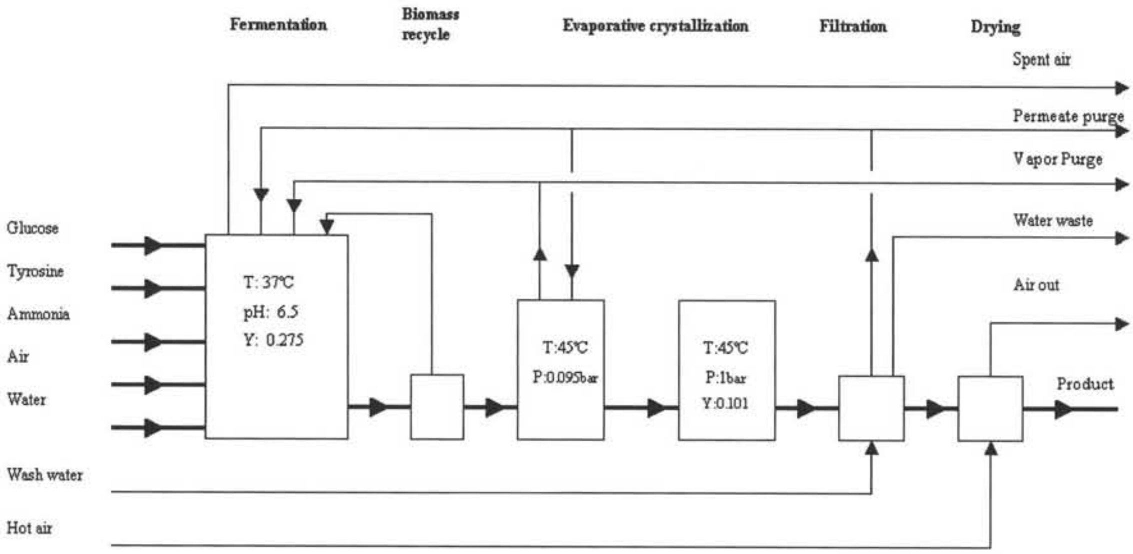

Figure A3.1 Imporved Process

Appendix 3.4 Economic Margin

Calculations for Net Cash Flow (NCF) are given as below:

NCF=Sales Income-Total Costs
Sales Income=Forecast Sales×Forecast Selling Price
Total Costs=Costs of Raw Material+Costs of Waste Treatment

Margin is obtained by dividing the NCF by the amount of the product. Detailed information for costs, sales, Net Cash Flow and margin is shown in the table below:

Table A3.12 Net Cash Flow and Margin for the process

	Price/ton [€/ton]	Quantity [ton]	Costs (k€)
Cost			
Glucose[1]	245.8	6400	1573
Tyrosine[2]	14749.3	85	1253
Mineralized Water[3]	0.1	14100	1.155
	Cost for Raw Materials A		2828
Waste Water/COD[4]	245.8	247.526	60.8
		Cost for wastewater Treatment B	60.8
	Total Cost C=A+B		2888
Sales			
L-phenylalanine[2]	8194	1000	8194
		Total Sales D	**8194**
		Net Cash Flow (D-C)	**5305**
		Margin (€/kg)	**5.3**

1. Price of glucose is obtained from International Starch Market Place Companies.htm, visited 03-08-2005
2. Price of tyrosine is obtained from Johannes Bongaerts, Metabolic Engineering for Microbial Production of Aromatic Amino Acids and Derived Compounds
3. Price of mineralized water is obtained from R.K. Sinnott, Coulson & Richardson's Chemical Engineering
4. Price of treatment for kg COD is obtained from Prof. Dr.ir. M.van Loosdrecht, TU Delft

The discounted cash flow Rate of return is a measure of the maximum rate that the project could pay and still break even by the end of the project life. The net cash flow in each year of the project is brought to its "present worth" at the start of the project by discounting it at some chosen interest rate (Sinnott, R.K., Coulson & Richardson's Chemical Engineering, Volume 6):

$$NPW_n = \frac{NFW_n}{(1+r)^n} \qquad \text{(A3.1)}$$

Where:

NPW_n: Net present worth of cash flow in year n

NFW_n: Future worth of the net cash flow in year n. (Estimated net cash flow in year n)

t: Life of plant, years

r: The discount Cash Flow Rate of Return (DCFROR)

Based on DCFRR and NCF, the total Net Present Worth of the project (NPW_{total}) can be obtained. And the maximum allowed investment can be calculated from the total Net Present Worth of the project.

$$NPW_{total} = \sum_{n=1}^{n=t} \frac{NFW_n}{(1+r)^n} \qquad \text{(A3.2)}$$

Where:

NPW_{total} : The total NPW of the project

Maximum allowed investment based on a certain DCFRR is the value at which the cumulative net present worth at the end of the project is zero, which means at the end of the project the investment will be returned. In this project, it is assumed that the plant (buildings, etc.) already exists and can be put into operation since the first year. Cumulated net present worth at the end of the project is the sum of the maximum allowed investment and the total NPW of the project. By knowing the cumulated net present worth at the end of the project is zero and the sum of the net present worth in each year, the maximum allowed investment can be determined, which is equal to NPW_{total}.

With the assumption that the life of the plant is 15 years and DCFRR is 10%, the total NPW_{total} and maximum allowed investment is calculated as below:

Table A.3.13 Total Net Present Worth and Maximum Allowed Investment in the Process

End of year	Future worth of Net Cash Flow in year n NFWn (M€)	1/(1+r)^n	Net Present Worth of cash flow in year n NPWn (€)	Maximum Allowed Investment (€)	Cumulative Net Present Worth NPW (€)
0	-	-	-	-40.4	0
1	5.31	0.9091	4.8	-	-40.4
2	5.31	0.8264	4.4	-	-35.5
3	5.31	0.7513	4.0	-	-31.1
4	5.31	0.6830	3.6	-	-27.2
5	5.31	0.6209	3.3	-	-23.5
6	5.31	0.5645	3.0	-	-20.2
7	5.31	0.5132	2.7	-	-17.2
8	5.31	0.4665	2.5	-	-14.5
9	5.31	0.4241	2.2	-	-12.0
10	5.31	0.3855	2.0	-	-9.8
11	5.31	0.3505	1.9	-	-7.8
12	5.31	0.3186	1.7	-	-5.9
13	5.31	0.2897	1.5	-	-4.2
14	5.31	0.2633	1.4	-	-2.7
15	5.31	0.2394	1.3		-1.3
Total	-	-	32.4	-	0.0

Reference: Coulson & Richardson's Chemical Engineering, Volume 6

APPENDIX 4.THERMODYNAMIC PROPERTIES

Appendix 4.1. Methods for Determining Thermodynamic Properties

Due to the lack of information for the thermodynamic properties of some of the compounds, the Joback method is used for their determination. This method consist on the use of the following equations using values reported for the groups that form the molecule of the component of interest to determine its properties in literature (Bruce E. Poling, (2001). The properties of gases and liquids)

Method of Joback

Vapor-liquid critical temperature, T_c, pressure, P_c, and volume, V_c

$$T_b = 198 + \sum_k N_k(tbk) \tag{A4.1}$$

$$T_{fp} = 122 + \sum_k N_k(tfpk) \tag{A4.2}$$

$$T_c(K) = T_b\left[0.584 + 0.965\left\{\sum_k N_k(tck)\right\} - \left\{\sum_k N_k(tck)\right\}^2\right]^{-1} \tag{A4.3}$$

$$P_c(bar) = \left[0.113 + 0.0032N_{atoms} - \sum_k N_k(pck)\right]^{-2} \tag{A4.4}$$

$$V_c(cm^3 mol^{-1}) = 17.5 + \sum_k N_k(vck) \tag{A4.5}$$

$$\Delta G_f^o(298.15K) = 53.88 + \sum_k N_k(gfk) \quad (\text{kJ mol}^{-1}) \tag{A4.6}$$

$$\Delta H_f^o(298.15K) = 68.29 + \sum_k N_k(hfk) \quad (\text{kJ mol}^{-1}) \tag{A4.7}$$

$$\Delta H_v = 15.30 + \sum_k N_k(hvk) \quad (\text{kJ mol}^{-1}) \tag{A4.8}$$

$$\Delta H_m = -0.88 + \sum_k N_k(hmk) \quad (\text{kJ mol}^{-1}) \tag{A4.9}$$

$$C_p^o(T) = \left[\sum_k N_k CpAk - 37.93\right] + \left[\sum_k N_k CpBk + 0.210\right] T + \left[\sum_k N_k CpCk - 3.91E-04\right] T^2 + \left[\sum_k N_k CpDk + 2.06E-07\right] T^3 \quad (\text{J mol}^{-1}\text{K}^{-1}) \qquad \text{(A4.10)}$$

$$A = \left[\sum_k N_k CpAk - 37.93\right] \qquad \text{(A4.10.1)}$$

$$B = \left[\sum_k N_k CpBk + 0.210\right] \qquad \text{(A4.10.2)}$$

$$C = \left[\sum_k N_k CpCk - 3.91E-04\right] \qquad \text{(A4.10.3)}$$

$$D = \left[\sum_k N_k CpDk + 2.06E-07\right] \qquad \text{(A4.10.4)}$$

For the calculation of the accentric factor the Ambrose-Walton method is used

$$\omega = \frac{\ln(P_c / 1.01325) + f(0)(T_{br})}{f(1)(T_{br})} \qquad \text{(A4.11)}$$

$$f(0) = \frac{(-5.97616T + 1.29874T^{1.5} - 0.60394T^{2.5} - 1.06841T^{5})}{T_{br}} \qquad \text{(A4.11.1)}$$

$$f(1) = \frac{(-5.03365T + 1.11505T^{1.5} - 5.41217T^{2.5} - 7.46628T^{5})}{T_{br}} \qquad \text{(A4.11.2)}$$

The critical compressibility factor is calculated with the following equation [BOD report]

$$Z_c = \frac{P_c V_c}{R T_c} \qquad \text{(A4.12)}$$

$$Z_c = 0.291 - 0.080\omega \qquad \text{(A4.13)}$$

Appendix 4.2. Joback values

An example of the calculation method is shown in Table A.4.1

Table A4.1 Joback method for the calculation of the thermodynamic properties of L- Phenylalanine, L-Tyrosine and Glucose

	N_k	*tfpk*	*tbk*	*tck*	*pck*	*vck*	*hfk*	*gfk*	*CpAk*	*CpBk*	*CpCk*	*CpDk*	*hvk*	*hmk*
Phenylalanine		*K*	*K*	*K*	*bar*	*cm^3/mol*	*kJ/ mol*	*kJ/ mol*	*J/mol/K*	*J/mol/K*	*J/mol/K*	*J/mol/K*	*kJ/mol*	*kJ/mol*
ACOH	0	82.83	76.34	0.024	0.0184	-25								
CH_2	1	11.27	22.88	0.0189	0	56	-20.64	8.42	-0.909	9.50E-02	-5.44E-05	1.19E-08	5.32E+02	6.19E+02
COOH	1	155.5	169.09	0.0791	0.0077	89	-426.72	-387.87	24.1	4.27E-02	8.04E-05	-6.87E-08	4.67E+03	2.64E+03
NH_2	1	66.89	73.23	0.0243	0.0109	38	-22.02	14.07	26.9	-4.12E-02	1.64E-04	-9.76E-08	2.58E+03	8.40E+02
CH= (ring)	5	8.13	26.73	0.0082	0.0011	41	2.09	11.3	-2.14	5.74E-02	-1.64E-06	-1.59E-08	6.08E+02	2.63E+02
C= (ring)	1	37.02	31.01	0.0143	0.0008	32	46.43	54.05	-8.25	1.01E-01	-1.42E-04	6.78E-08	7.31E+02	5.72E+02
CH	1	12.64	21.74	0.0164	0.002	41	29.89	58.36	-23	2.04E-01	-2.65E-04	1.20E-07	4.04E+02	1.79E+02
$\sum_k N_k F_k$		323.97	451.6	0.194	0.0269	461	-382.61	-196.47	8.141	0.6885	-0.00023	-4.61E-08	11954	6166
Tyrosine														
ACOH	1	82.83	76.34	0.024	0.0184	-25	-221.65	-197.37	-2.81	1.11E-01	-1.16E-04	4.94E-08	2.99E+03	1.07E+03
CH_2	1	11.27	22.88	0.0189	0	56	-20.64	8.42	-0.909	9.50E-02	-5.44E-05	1.19E-08	5.32E+02	6.19E+02
COOH	1	155.5	169.09	0.0791	0.0077	89	-426.72	-387.87	24.1	4.27E-02	8.04E-05	-6.87E-08	4.67E+03	2.64E+03
NH_2	1	66.89	73.23	0.0243	0.0109	38	-22.02	14.07	26.9	-4.12E-02	1.64E-04	-9.76E-08	2.58E+03	8.40E+02
CH= (ring)	4	8.13	26.73	0.0082	0.0011	41	2.09	11.3	-2.14	5.74E-02	-1.64E-06	-1.59E-08	6.08E+02	2.63E+02
C= (ring)	2	37.02	31.01	0.0143	0.0008	32	46.43	54.05	-8.25	1.01E-01	-1.42E-04	6.78E-08	7.31E+02	5.72E+02
CH	1	12.64	21.74	0.0164	0.002	41	29.89	58.36	-23	2.04E-01	-2.65E-04	1.20E-07	4.04E+02	1.79E+02
$\sum_k N_k F_k$		435.69	532.22	0.2241	0.045	427	-559.92	-351.09	-0.779	0.8431	-0.00048	8.7E-08	15064	7548
Glucose														
OH	5	44.45	92.88	0.0741	0.0112	28	-208.04	-189.2	25.7	-6.91E-02	1.77E-04	-9.88E-08	4021	575
CH2	1	11.27	22.88	0.0189	0	56	-20.64	8.42	-0.909	9.50E-02	-5.44E-05	1.19E-08	5.32E+02	6.19E+02
CH	4	12.64	21.74	0.0164	0.002	41	29.89	58.36	-23	2.04E-01	-2.65E-04	1.20E-07	4.04E+02	1.79E+02
CHO	1	36.9	72.2	0.0379	0.003	82	-162.03	-143.48	30.9	-3.36E-02	1.60E-04	9.88E-08	2173	764
$\sum_k N_k F_k$		320.98	646.44	0.4929	0.067	442	-1103.31	-847.62	66.491	0.5319	-6.9E-05	9.67E-08	24426	4974

Table A4.2 Calculated properties of L-Phenylalanine, L-Tyrosine and Glucose

		L-Phenylalanine	L-Tyrosine	Glucose
Properties	**Unit**	**Value**	**Value**	**Value**
Tb	K	649.6	730.22	844.44
Tfp	K	445.97	557.69	442.98
Tc	K	885.5275678	973.5803372	1033.968379
Pc	bar	39.20939739	47.69390434	66.31370094
Vc	cm3/mol	478.5	444.5	459.5
det H_f^o	*kJ/ mol*	-314.32	-491.63	-1035.02
det G_f^o	*kJ/ mol*	-142.59	-297.21	-793.74
det Hv	*kJ/ mol*	11969.3	15079.3	24441.3
det Hm	*kJ/ mol*	6165.12	7547.12	4973.12
R	*J/(molK)*	8.3145	8.3145	8.3145
Zc	*[-]*	0.25482022	0.2618947	0.35444217
Tr	*[-]*	0.733574	0.75003569	0.81669809
t	*[-]*	0.266426	0.24996431	0.18330191
f(0)		-1.959126088	-1.801818242	-1.227414828
f(1)		-1.903110243	-1.726901139	-1.119839423
w		-1.58916531	-1.930312122	-3.475735811
A		-29.789	-38.709	28.561
B		0.8985	1.0531	0.7419
C		-0.0006162	-0.00087256	-0.0004604
D		1.599E-07	0.000000293	3.027E-07
T	K	298.15	298.15	298.15
Cp	*J/mol/K*	187.5605695	205.4734674	216.8545851
Zc2	[-]	0.418133225	0.44542497	0.569058865

Appendix 4.2. Pathway of Aromatic Amino Acid Synthesis[1]

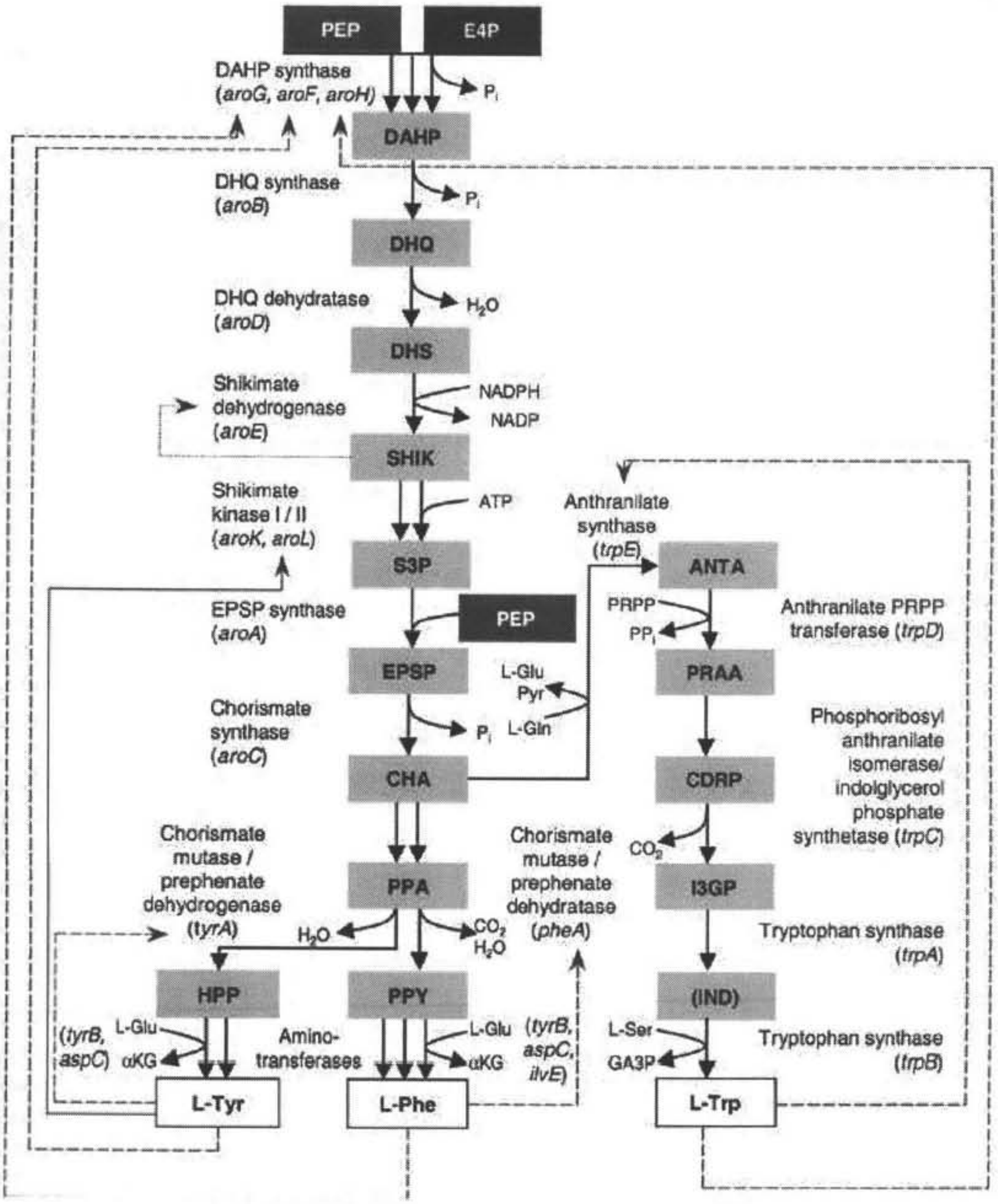

Figure A4.1 Metabolism of L-Phenylalanine Production

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[1] Taken from the article by Bongaerts et al, 2001. Metabolic Engineering for Microbial Production of Aromatic Amino Acids and Derived Compounds

Appendix 4.3 Solubility and Metastability Line for L-Phenylalanine Crystals

The Solubility of the anhydrate and monohydrate crystals are shown in the following table. The constants for the calculation are taken from Mohan et al, 2001

$T := 0,1..60$

$C\_metastability(T) := 0.00887 \cdot T^2 - 0.0598 \cdot T + 29.681$

$a1 := 2.043$	$a2 := 1.949$
$b1 := 0.02394$	$b2 := 0.005169$
$c1 := 0.0004211$	$c2 := 0.00099$

$Cphe\_anhydrate(T) := \left(a1 + b1 \cdot T + c1 \cdot T^2\right) \cdot 10$

$Cphe\_monohydrate(T) := \left(a2 + b2 \cdot T + c2 \cdot T^2\right) \cdot 10$

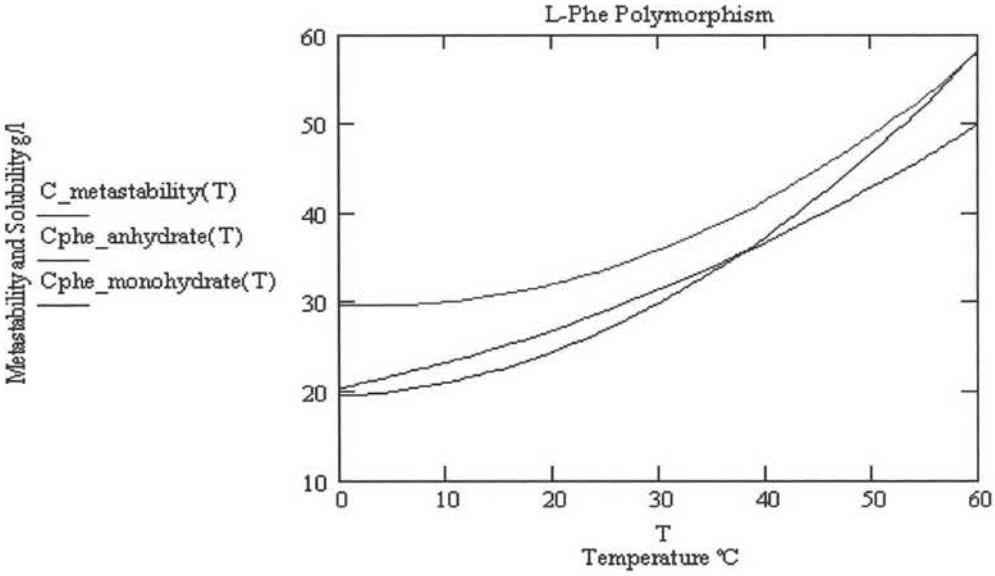


Figure A4.2 Production rate as a function of Concentration of L-Phenylalanine

Appendix 4.4. Phenylalanine Production

Kiphe := 20	g_phe l^{-1}
qphemax := 0.0756	g_phe gX^{-1} h^{-1}
qmtyr := 0.00245	g_tyr $gX^{-1}h^{-1}$
Yx_tyr := 35.29	gX g_tyr^{-1}
qmglu := 0.05	gX g_glu^{-1} h^{-1}
Yx_glu := 0.51	gX g_glu^{-1}
Yphe_glu := 0.275	g_phe g_glu^{-1}

Assuming that the system is at steady state so no growth is observed μ=0

$\mu := 0$

Cphe := 0,001 .. 40

$$\mathrm{qphe(Cphe)} := \mathrm{qphemax}\cdot\frac{\mathrm{Kiphe}}{\mathrm{Kiphe} + \mathrm{Cphe}}$$

$$\mathrm{qtyr} := \mathrm{qmtyr} + \frac{\mu}{\mathrm{Yx\_tyr}}$$

$$\mathrm{qglu(Cphe)} := \mathrm{qmglu} + \frac{\mu}{\mathrm{Yx\_glu}} + \frac{\mathrm{qphe(Cphe)}}{\mathrm{Yphe\_glu}}$$

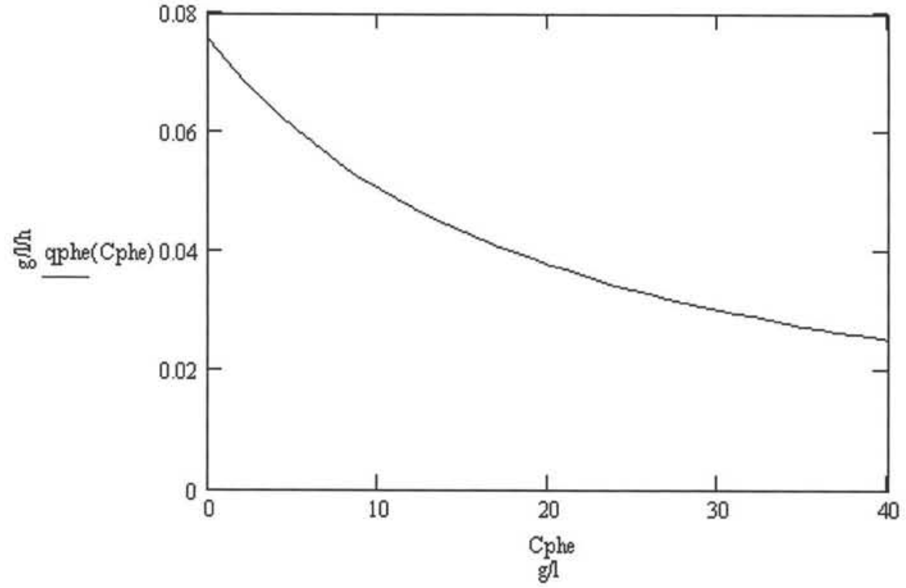


Figure A4.3 Phenylalanine Production related with its concentration

Appendix 4.5 Stoichiometry Calculations

Stoichiometry calculations are done in MathCad and the related file (FermentorPhaseCalculations.mcd) can be found in the CD. In this part the results of the calculations will be presented to give an insight to fermentor design.

- Growth Reaction

$-0.268\ C_6H_{12}O_6 - 0.2\ NH_4^+ - 0.556\ O_2 + 1\ CH_{1.8}O_{0.5}N_{0.2} + 0.2\ H^+ + 0.606\ CO_2 + 1.006\ H_2O + 252.805$ kJ/mol

- Production Reaction

$-\ 3.334\ C_6H_{12}O_6 - 1NH_4^+ -10.005\ O_2 + C_9H_{11}NO_2 + 11.005\ CO_2 + H^+ + 16.005\ H_2O^- + 4882$ kJ/mol

- Catabolic Reaction

$-1\ C_6H_{12}O_6 - 6\ O_2 + 6\ CO_2 + 6\ H_2O$

- Growth Rate, μ

$$\mu\left(C_{Tyr}\right) := \mu_{max} \cdot \frac{C_{Tyr}}{C_{Tyr} + K_{S\_Tyr}}$$

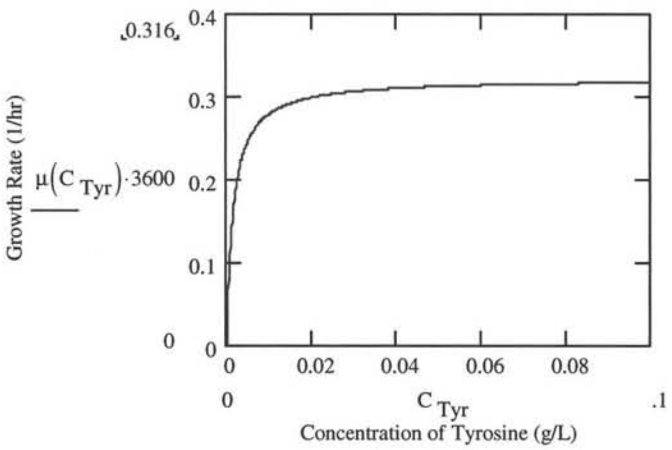


Figure A4..4 Growth Rate

- Phenylalanine Production, q_{phe}

$$q_{Phe}(C_{Phe}) := q_{max\_Phe} \cdot \frac{K_{I\_Phe}}{K_{I\_Phe} + C_{Phe}}$$

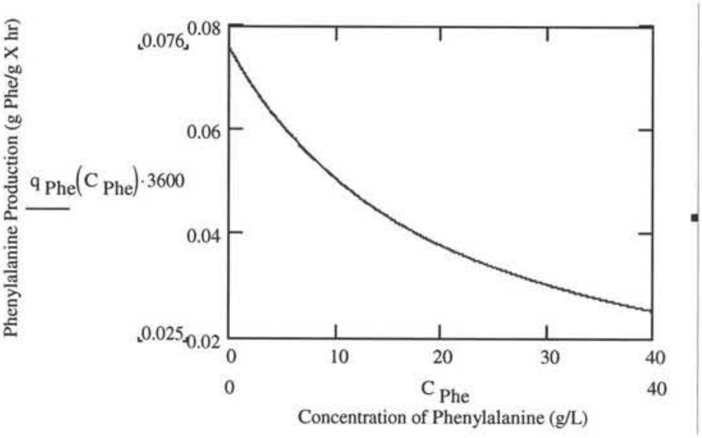


Figure A4..5 Phenylalanine Production

- Tyrosine Consumption

$$q_{Tyr}(\mu, C_{Tyr}) := \frac{-\mu(C_{Tyr})}{Y_{X\_Tyr}} + q_{m\_Tyr}$$

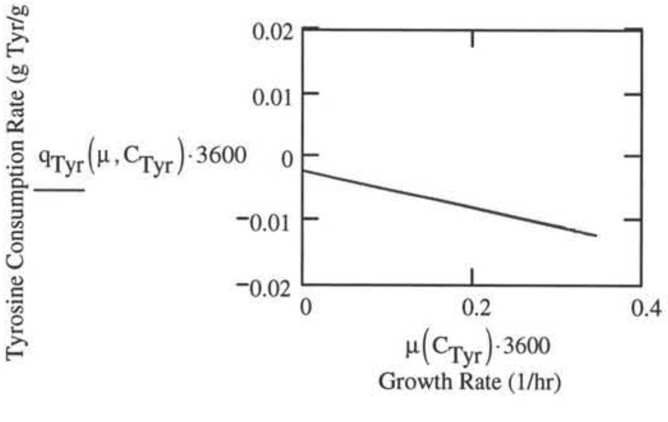


Figure A4..6 Tyrosine consumption related with growth rate

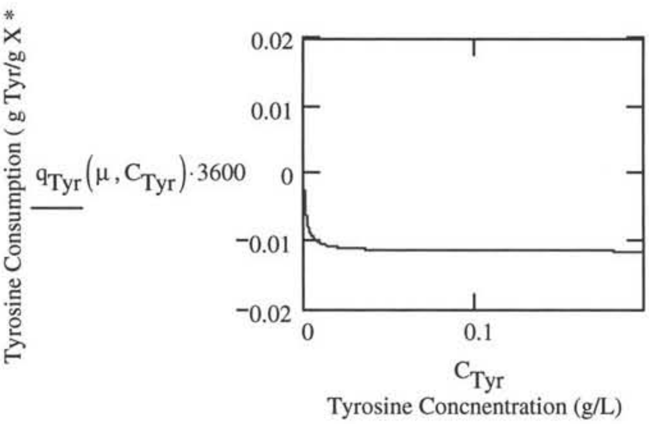


Figure A4..7 Tyrosine consumption related with tyrosine concentration

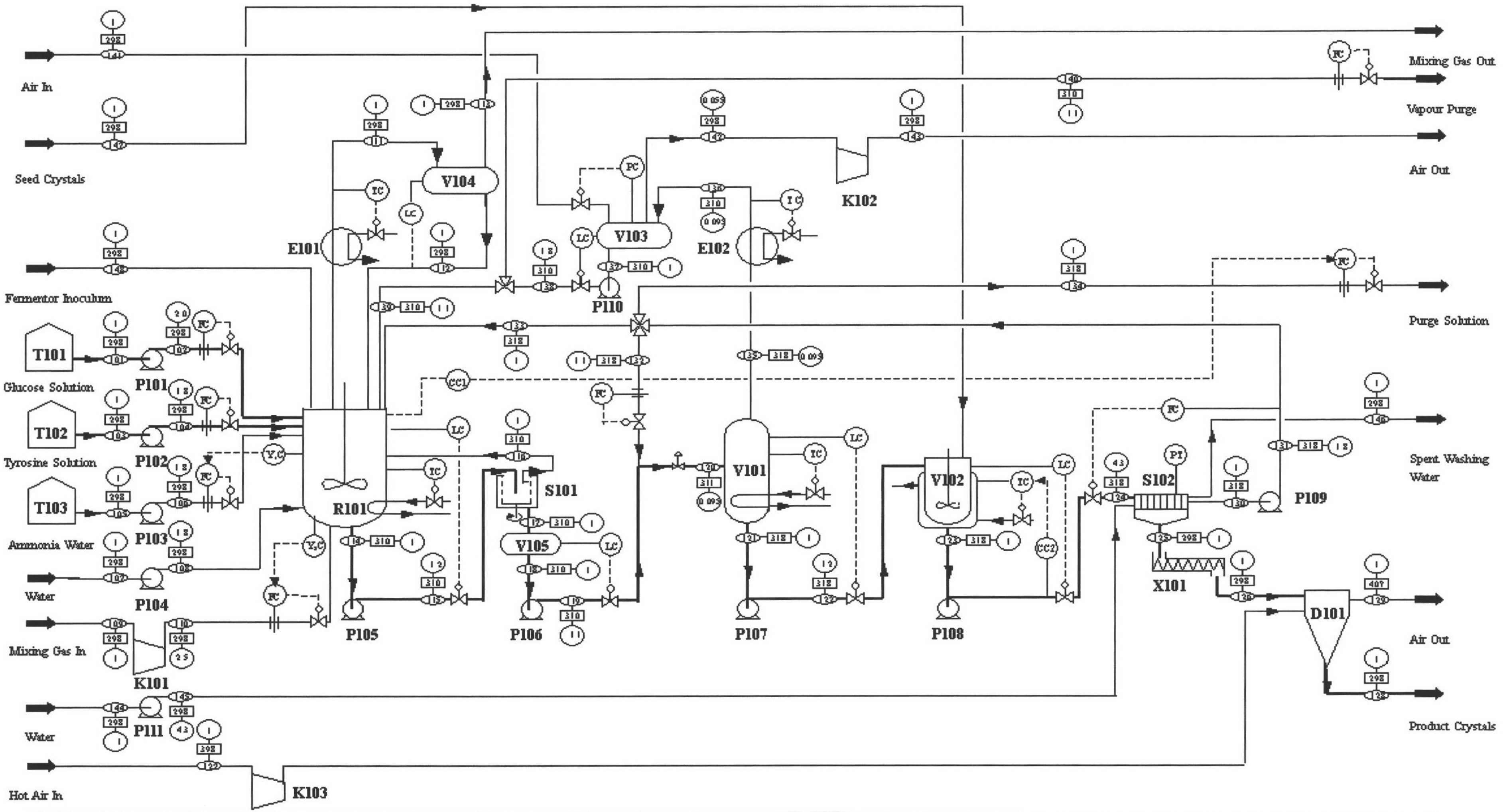
These stoichiometry calculations are used in mass balance calculations of both batch and production phases.

Appendix 5 PROCESS STRUCTURE AND DESCRIPTION

Appendix 5.1 Process flow schemes

This appendix includes the process flow schemes, batch operations, and the process stream summary

Appendix 5.1.1 Main Process Flow Scheme



Process Equipment Summery

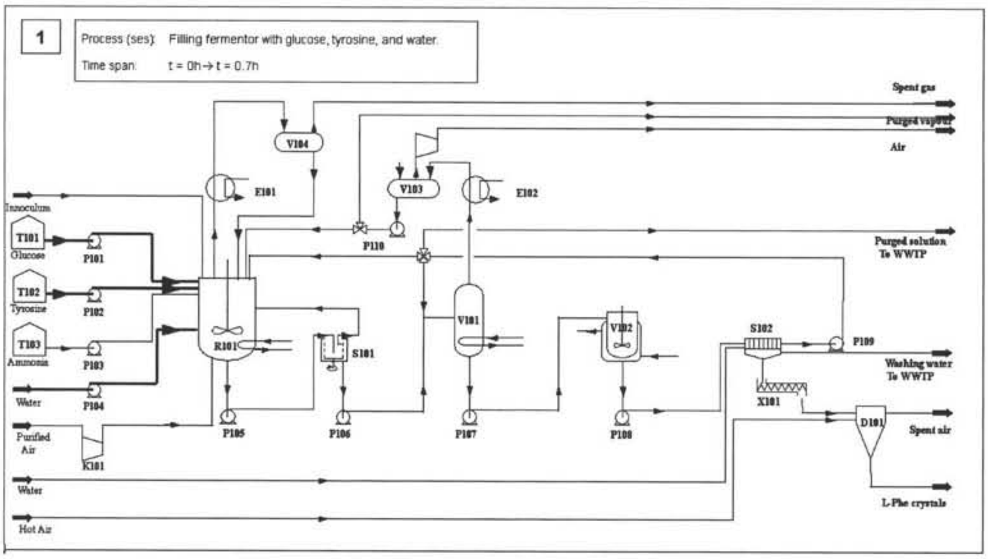
D101: Dryer
E101: Water Removal V104
E102: Condensor V103
P101: Feeding Glucose
P102: Feeding Tyrosine
P103: Feeding Ammonia
P104: Filling R101
P105: R101 Withdraw
P106: V105 Withdraw
P107: V101 Withdraw
P108: V102 Withdraw
P109: Effluent Withdraw
P110: V103 Withdraw
P111: Feeding Wash Water
R101: Fermentor
S101: Centrifuge
S102: Plate-frame filter
T101: Storage glucose
T102: Storage tyrosine
T103: Storage ammonia
V101: Evaporator
V102: Crystallizer
V103: Vessel E102
V104: Vessel E101
V105: Vessel S101
X101: Belt

Designers

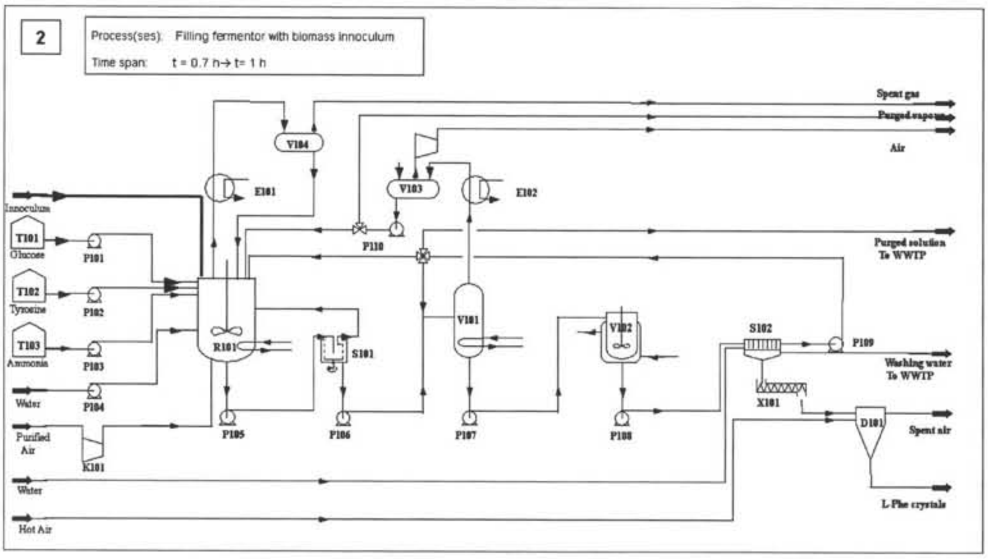
Juan Felipe Solarte
Mehmet Hikmet Ucisik
Huiling Yang
Jing Nie

Process Flow Scheme-Improved Solution
Project: Design of an Integrated Fermentation-Crystallization Process for Production of L-phenylalanine
Proj. ID Number: CPD 3324
Completion Date: 16th October 2005
⬭ Stream number ▭ Temp. (K) ○ Pressure (Bara)

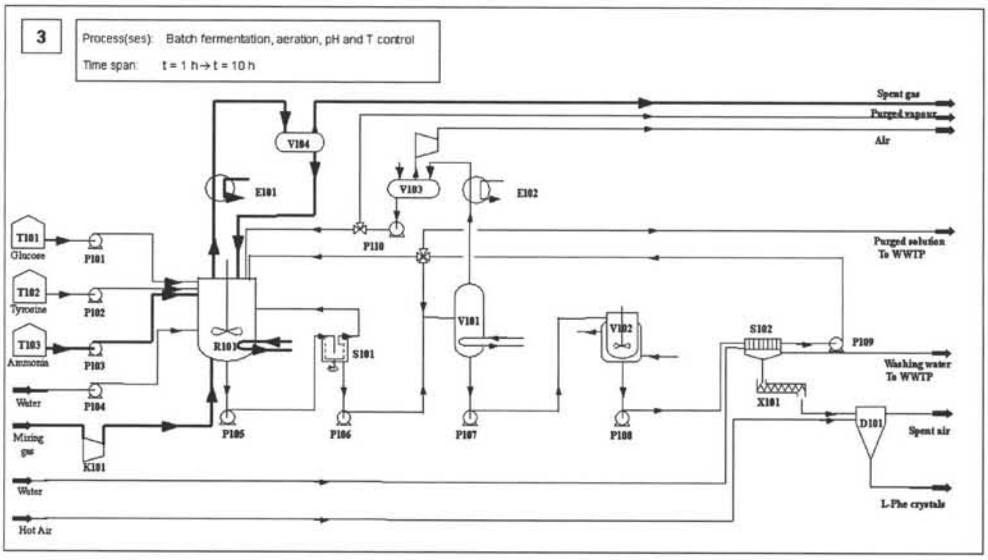
Appendix 5.1.2 Batch operation



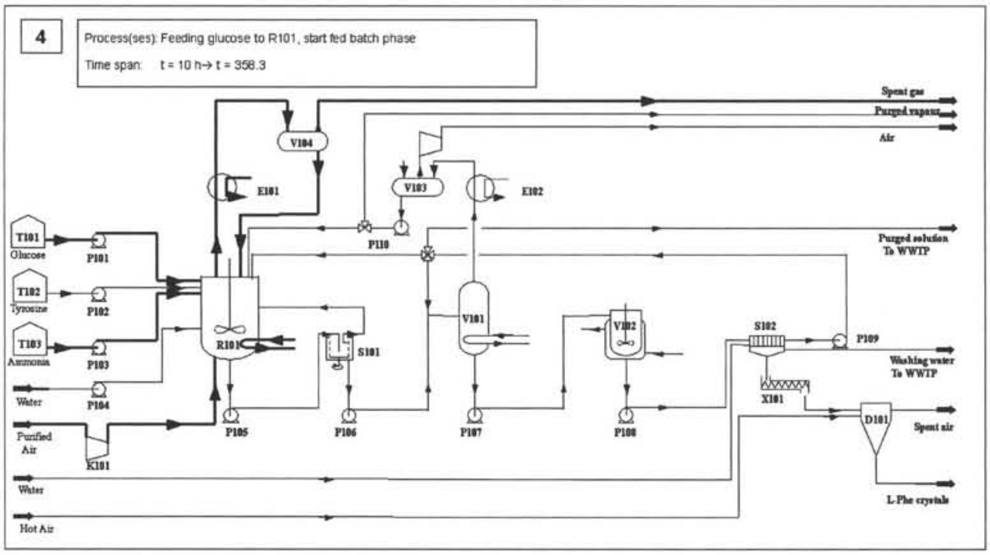
Process (ses): Filling fermentor with glucose, tyrosine, and water.
Time span: t = 0h→t = 0.7h



Process(ses): Filling fermentor with biomass innoculum
Time span: t = 0.7 h→ t= 1 h



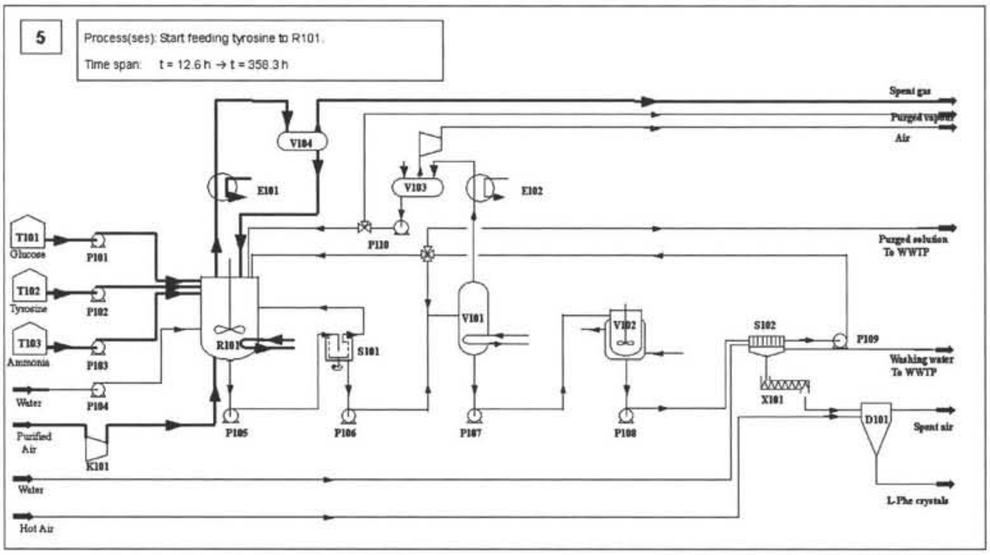
Process(ses): Batch fermentation, aeration, pH and T control
Time span: t = 1 h→t = 10 h



4

Process(ses): Feeding glucose to R101, start fed batch phase

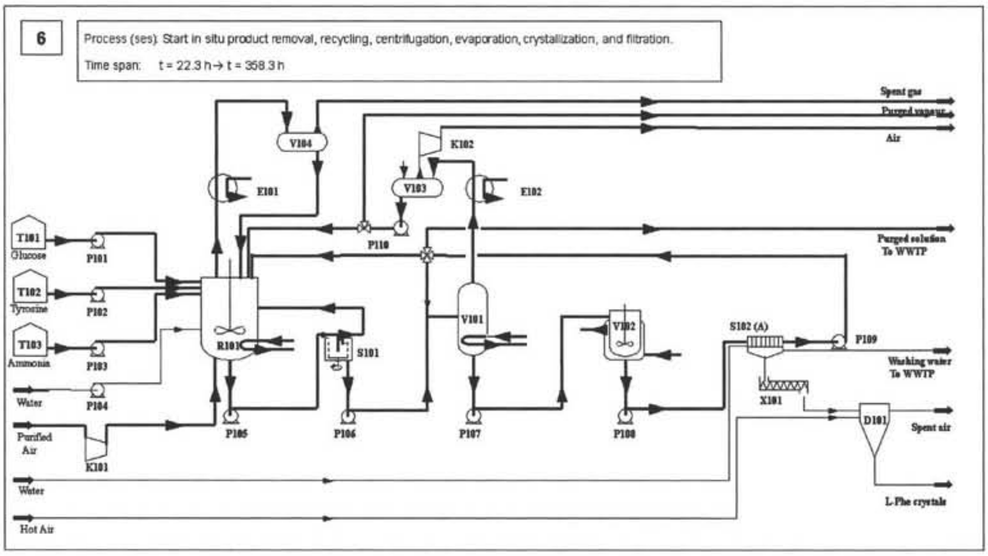
Time span: t = 10 h→ t = 358.3



5

Process(ses): Start feeding tyrosine to R101.

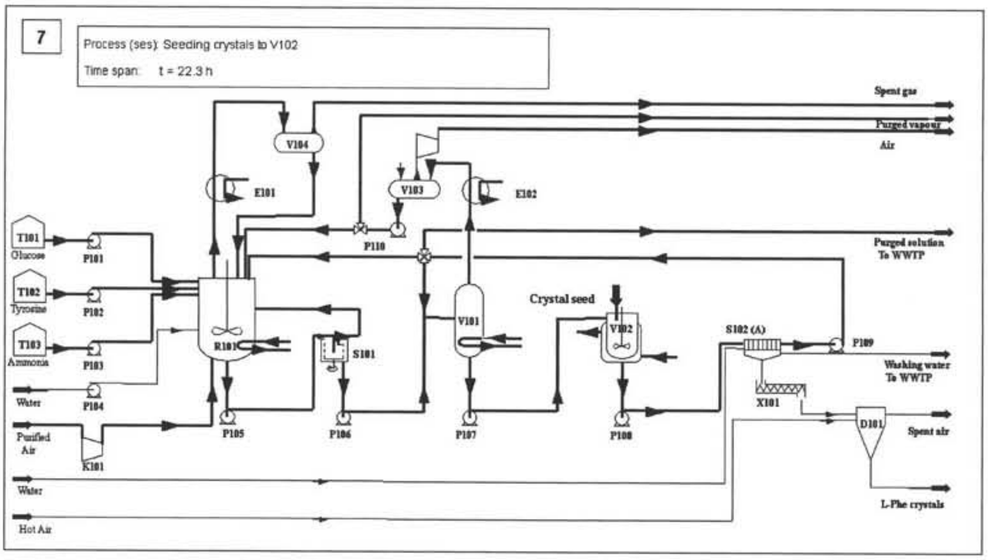
Time span: t = 12.6 h → t = 358.3 h



6

Process (ses): Start in situ product removal, recycling, centrifugation, evaporation, crystallization, and filtration.

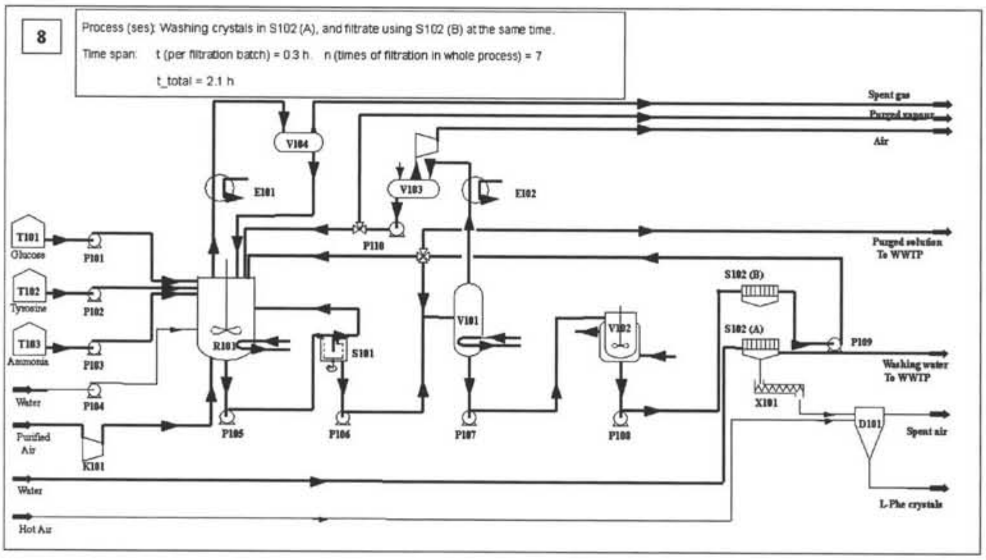
Time span: t = 22.3 h→ t = 358.3 h



7

Process (ses): Seeding crystals to V102

Time span: t = 22.3 h

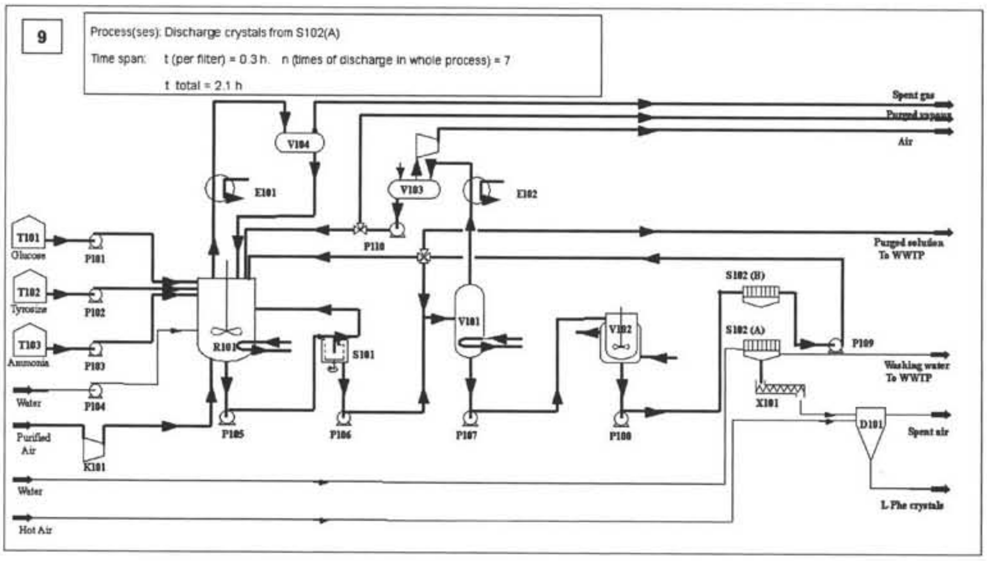


8

Process (ses): Washing crystals in S102 (A), and filtrate using S102 (B) at the same time.

Time span: t (per filtration batch) = 0.3 h. n (times of filtration in whole process) = 7

t_total = 2.1 h

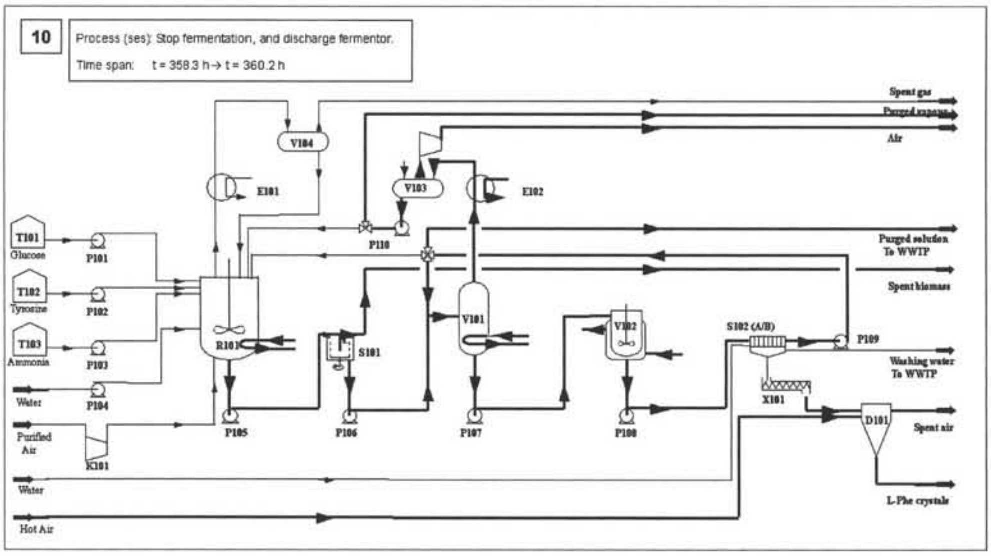


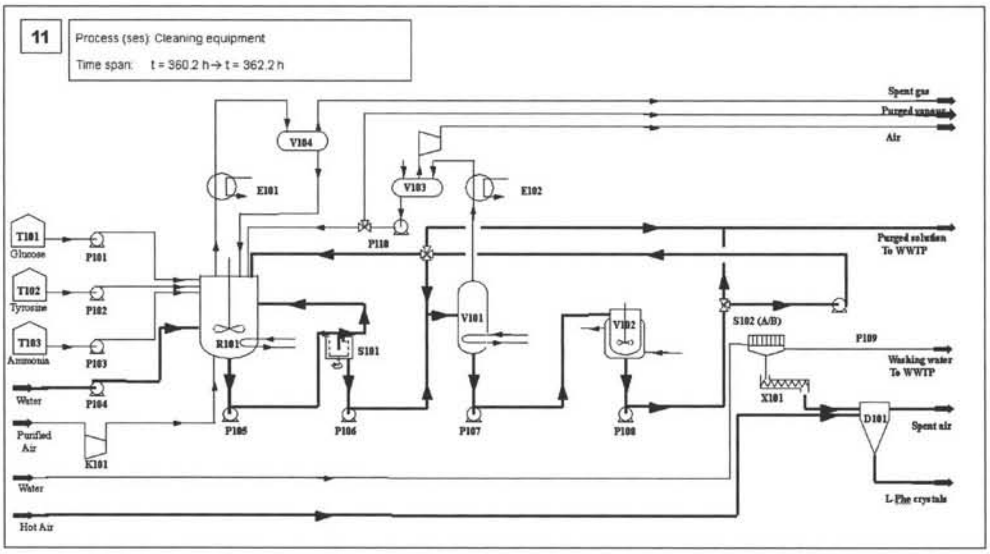
9

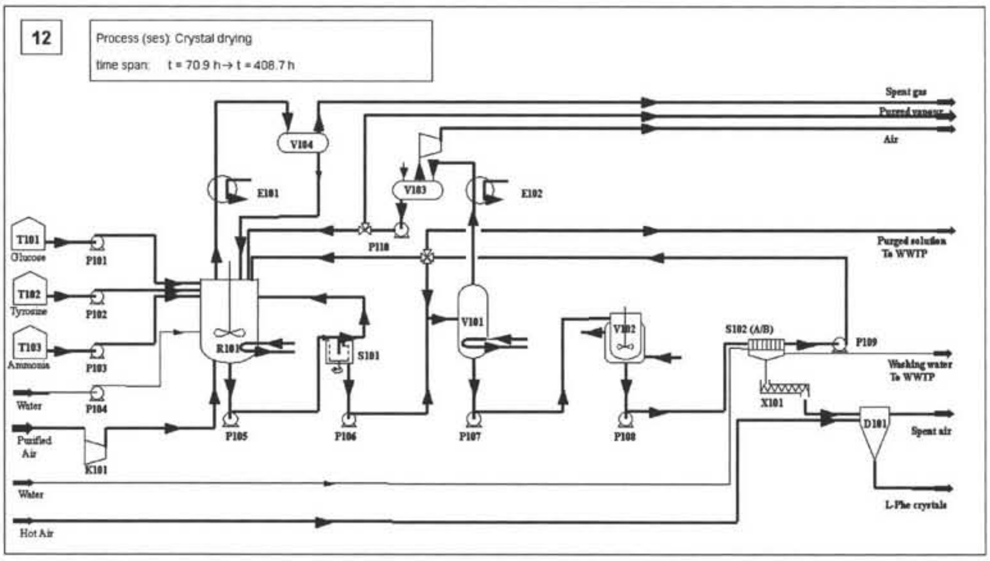
Process(ses): Discharge crystals from S102(A)

Time span: t (per filter) = 0.3 h. n (times of discharge in whole process) = 7

t total = 2.1 h


10
Process (ses): Stop fermentation, and discharge fermentor.
Time span: t = 358.3 h → t = 360.2 h
Spent gas
Purged vapour
Air
V104
E101
V103
E102
P110
Purged solution To WWTP
Spent biomass
T101
Glucose
P101
T102
Tyrosine
P102
T103
Ammonia
P103
Water
P104
Purified Air
K101
R101
S101
V101
V102
S102 (A/B)
P109
Washing water To WWTP
X101
D101
Spent air
P105
P106
P107
P108
Water
L-Phe crystals
Hot Air


11
Process (ses): Cleaning equipment
Time span: t = 360.2 h → t = 362.2 h
Spent gas
Purged vapour
Air
V104
E101
V103
E102
P110
Purged solution To WWTP
T101
Glucose
P101
T102
Tyrosine
P102
T103
Ammonia
P103
Water
P104
Purified Air
K101
R101
S101
V101
V102
S102 (A/B)
P109
Washing water To WWTP
X101
D101
Spent air
P105
P106
P107
P108
Water
L-Phe crystals
Hot Air


12
Process (ses): Crystal drying
time span: t = 70.9 h → t = 408.7 h
Spent gas
Purged vapour
Air
V104
E101
V103
E102
P110
Purged solution To WWTP
T101
Glucose
P101
T102
Tyrosine
P102
T103
Ammonia
P103
Water
P104
Purified Air
K101
R101
S101
V101
V102
S102 (A/B)
P109
Washing water To WWTP
X101
D101
Spent air
P105
P106
P107
P108
Water
L-Phe crystals
Hot Air

Appendix 5.2 Batch Cycle Diagrams

The table below shows the relevant equipment, streams and process time for the batch operations. The process flow sheets for batch operation and batch cycle time are set up base on the times given in this table.

Table A5.1: Overview batch operations

Description	From	To	Pump	Stream(s)	Time (h)
Filling glucose to Fermentor (1)	T101	R101	P101	<101>/<102>	0.17
Filling tyrosine to Fermentor (1)	T102	R101	P102	<103>/<104>	0.06
Filling water to Fermentor		R101	P104	<107>/<108>	0.7
Inoculum to Fermentor		R101		<148>	0.3
Feeding ammonia to fermentor	T103	R101		<105>/<106>	357.3
Aeration fermentor		R101	K101	<109>/<110>	357.3
Feeding glucose to fermentor (2)	T101	R101	P101	<101>/<102>	348.3
Feeding tyrosine to fermentor (2)	T102	R101	P102	<103>/<104>	345.7
Product in situ removal	R101				336
Filtration crystals (3)	V102	P108	S102	<123>/<124>	48
Washing crystal (3)		P111	S102	<144>/<145>	0.3
Discharging crystal (3)	S102	V (4)		<125>	0.3
Drying crystal	V (4)	D101	X101	<126>	337.9
Discharge fermentor	R101	To burn		<124>/<115>/<116>	1.9
Cleaning fermentor		WWTP	P104 P105 S105 P106 P107 P108	<114>/<115>/<116> <117>/<118>/<119> <120>/<121><122> <123>/<124>/<132> <133>	2

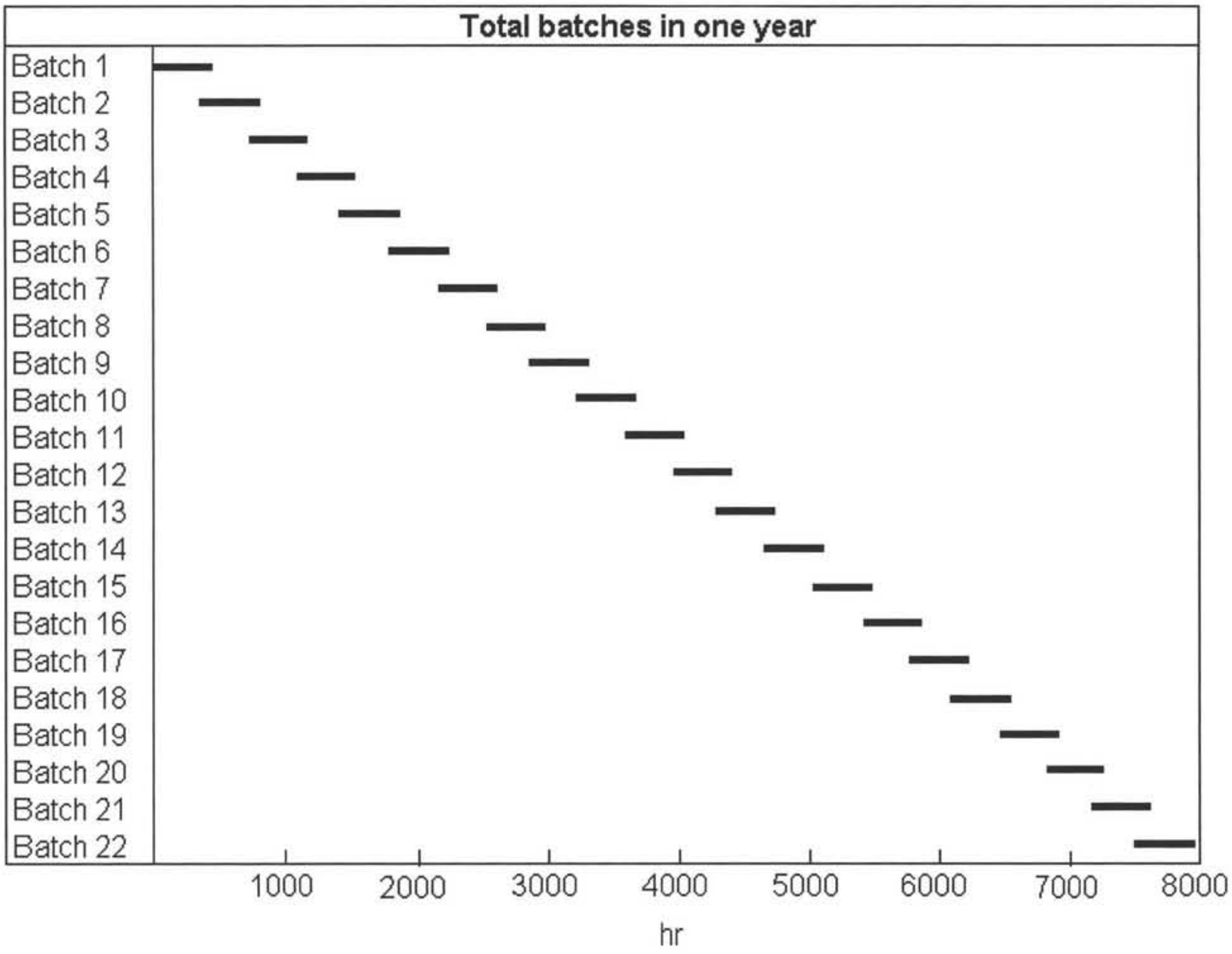


Figure A5.1 Batch process cycle diagram for one year of production.

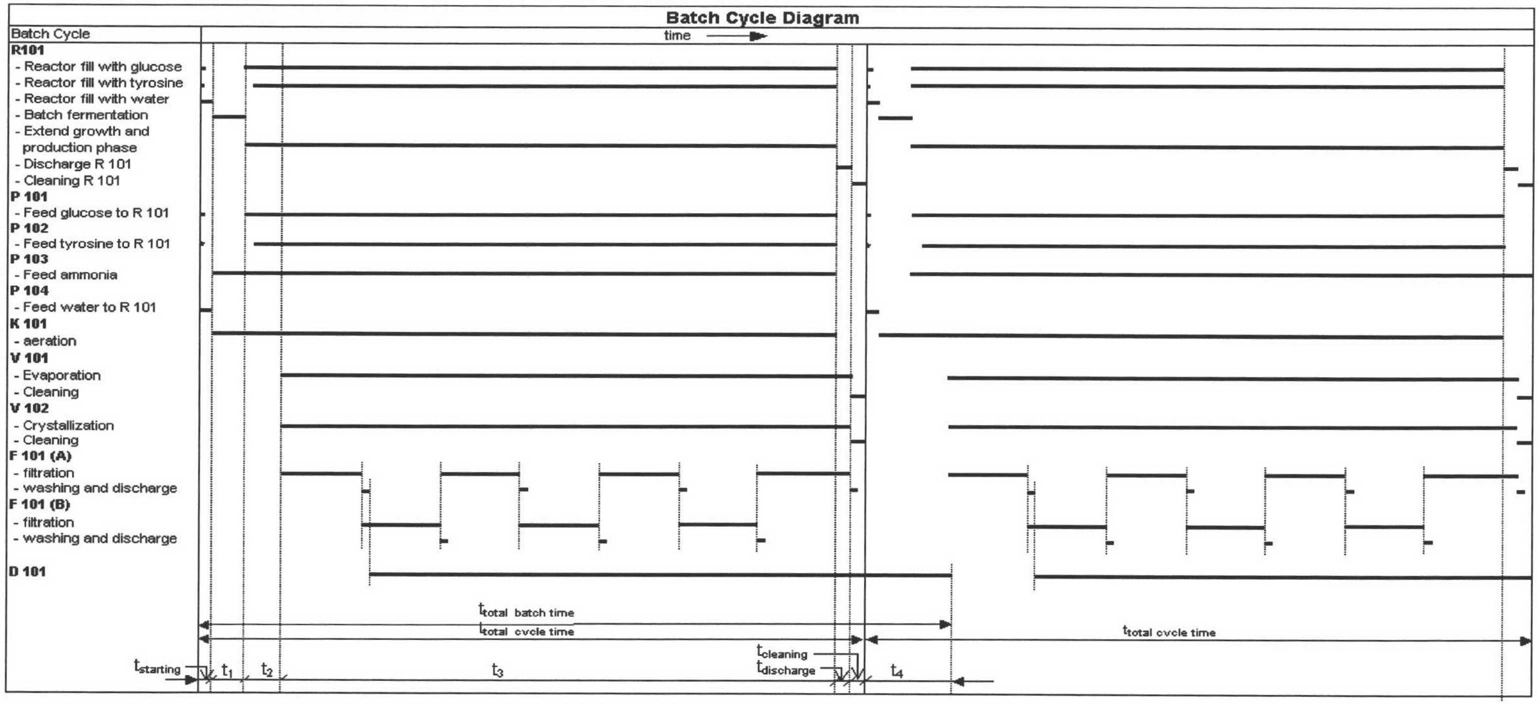


Figure A5.2 Batch cycle Diagram shows detailed batch cycle of one make span and two cycle times.

Table A5.2 Detailed description of Figure A5.2

Cycle times		Unit	Remarks
$t_{starting}$	1.0	[h]	Time for feeding water, glucose, tyrosine, and inoculum
t_1	9.0	[h]	Biomass growth in batch phase
t_2	12.3	[h]	From to ending of batch to the starting of steady state production phase
t_3	336.0	[h]	Steady state production phase
$t_{discharge}$	1.9	[h]	Discharge fermentor after fermentation finish
$t_{cleaning}$	2.0	[h]	Assume the same cleaning time for all the equipment
t_4	48.5	[h]	Drying time needed after cleaning equipment
$t_{total\ cycle\ time}$	362.2	[h]	
$t_{total\ make\ span}$	408.7	[h]	

Appendix 5.3 Process Stream Summery

Appendix 5.3.1 Process Stream Summary for Continuous Production Phase

STREAM Nr. :		101/102	IN	103/104	IN	105/106	IN	109/110	IN	113	OUT
Name :		Feed Glucose to R101		Feed tyrosine to R 101		Ammonia to R101		Mixing gas to R101		Off gas of R101	
COMP	MW	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr
Glucose	180.160	965.83	5.3610	0	0.0000	0	0.0000	0	0.0000	0	0.0000
Tyrosine	181.190	0	0.0000	8.698	0.0480	0	0.0000	0	0.0000	0	0.0000
L-Phe	165.190	0	0.0000	0	0.0000	0	0.0000	0	0.0000	0	0.0000
Acetate	59.000	0	0.0000	0	0.0000	0	0.0000	0	0.0000	0	0.0000
Biomass	24.600	0	0.0000	0	0.0000	0	0.0000	0	0.0000	0	0.0000
NH4	18.000	0	0.0000	0	0.0000	18.088	1.0049	0	0.0000	0	0.0000
H2O	18.000	413.927	22.9959	339.203	18.8446	54.265	3.0147	0	0.0000	0	0.0000
O2	32.000	0.00	0.0000	0.00	0.0000	0.00	0.0000	488.25	15.2579	0.10	0.0032
CO2	44.000	0.00	0.0000	0.00	0.0000	0.00	0.0000	1.17	0.0265	713.43	16.2144
N2	28.000	0.00	0.0000	0.00	0.0000	0.00	0.0000	1633.00	58.3214	1633.00	58.3214
Total		1379.76	28.3569	347.90	18.8926	72.35	4.0196	2122.42	73.6058	2346.54	74.5390
Enthalpy	kW	-3086		-1273		-215		-3		-1772	
Phase		L		L		L		G		G	
Press.	Bara	1.0/2		1.0/1.8		20.8		1.0/2.5		1.0	
Temp	oC	25		25		50.0		25		25	

STREAM Nr. :		114/115	IN	116		117/118/119		120 = 132+119	Cal	121	
Name :		Stream leaving R101		Biomass recycle to R101		Stream leaving S101		Stream to V101		Liquid stream leaving V101	
COMP	MW	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr
Glucose	180.160	313.422	1.7397	52.237	0.2899	261.185	1.4497	481.40	2.6721	481.398	2.6721
Tyrosine	181.190	0.0002026	0.0000	0.00003377	0.0000	0.0001688	0.0000	0.00	0.0000	0.0003112	0.0000
L-Phe	165.190	940.266	5.6920	156.711	0.9487	783.555	4.7434	1330.37	8.0536	1330	8.0513
Acetate	59.000	138.967	2.3554	23.161	0.3926	115.806	1.9628	213.45	3.6177	213.446	3.6177
Biomass	24.600	1881	76.4634	1881	76.4634	0	0.0000	0.00	0.0000	0	0.0000
NH4	18.000	3.197	0.1776	0.533	0.0296	2.664	0.1480	4.91	0.2728	4.91	0.2728
H2O	18.000	59410	3300.5556	8334	463.0000	51070	2837.2222	63970.00	3553.8889	28210	1567.2222
O2	32.000	0.17	0.0054	0.03	0.0009	0.14	0.0045	0.17	0.0053	0.06	0.0019
CO2	44.000	0.03	0.0006	0.00	0.0001	0.02	0.0005	0.03	0.0006	0.01	0.0002
N2	28.000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Total		62687.05	3386.9897	10447.68	541.1252	52233.38	2845.5311	66000.32	3568.5110	30239.82	1581.8382
Enthalpy	kW	-224106		-33142		-190942		-3		-106511	
Phase		L/S		L/S		L		L		L	
Press.	Bara	1.0/1.2		1.0		1.0/1.0/1.1		0.095		1.0	
Temp	oC	37		37		37		38		45	

STREAM Nr. :		122		123		124		130/131		132	
Name :		Stream to V102		Stream leaving V102		Stream to S102 (A/B)		Permeate leaving S102 (A/B)		Permeate recycle to V101	
COMP	MW	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr
Glucose	180.160	481.398	2.6721	481.612	2.6732	481.612	2.6732	480.172	2.6653	220.213	1.2223
Tyrosine	181.190	0.0003112	0.0000	0.0003114	0.0000	0.0003114	0.0000	0.0003105	0.0000	0.0001424	0.0000
L-Phe	165.190	1330	8.0513	1331	8.0574	1331	8.0574	1192	7.2159	546.817	3.3102
Acetate	59.000	213.446	3.6177	213.54	3.6193	213.54	3.6193	212.902	3.6085	97.64	1.6549
Biomass	24.600	0	0.0000	0	0.0000	0	0.0000	0	0.0000	0	0.0000
NH4	18.000	4.91	0.2728	4.912	0.2729	4.912	0.2729	4.898	0.2721	2.246	0.1248
H2O	18.000	28210	1567.2222	28200	1566.6667	28200	1566.6667	28120	1562.2222	12900	716.6667
O2	32.000	0.06	0.0019	0.06	0.0019	0.06	0.0019	0.06	0.0019	0.03	0.0009
CO2	44.000	0.01	0.0002	0.01	0.0002	0.01	0.0002	0.01	0.0002	0.00	0.0001
N2	28.000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Total		30239.82	1581.8382	30231.13	1581.2916	30231.13	1581.2916	30010.04	1575.9861	13766.95	722.9799
Enthalpy	kW	-106511		-106481		-106481		-106119		-48681	
Phase		L		L/S		L/S		L		L	
Press.	Bara	1.2		1.0		4.3		1.0/1.8		1.1	
Temp	oC	45		45		45		45		45	

STREAM Nr. :		133		134	OUT	135		136		137/138	
	Name :	Permeate recycle to R101		Permeate purge stream		Vapour leaving V101		After E102		Stream leaving V103	
COMP	MW	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr
Glucose	180.160	253.791	1.4087	6.167	0.0342	0	0.0000	0	0.0000	0	0.0000
Tyrosine	181.190	0.0001641	0.0000	0.000003988	0.0000	0	0.0000	0	0.0000	0	0.0000
L-Phe	165.190	630.195	3.8150	15.314	0.0927	0	0.0000	0	0.0000	0	0.0000
Acetate	59.000	112.528	1.9073	2.735	0.0464	0	0.0000	0	0.0000	0	0.0000
Biomass	24.600	0	0.0000	0	0.0000	0	0.0000	0	0.0000	0	0.0000
NH4	18.000	2.589	0.1438	0.063	0.0035	0	0.0000	0	0.0000	0	0.0000
H2O	18.000	14860	825.5556	361.179	20.0655	35760	1986.6667	35760	1986.6667	35760	1986.6667
O2	32.000	0.03	0.0010	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
CO2	44.000	0.00	0.0001	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N2	28.000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Total		15859.14	832.8314	385.46	20.2423	35760.00	1986.6667	35760.00	1986.6667	35760.00	1986.6667
Enthalpy	kW	-56080		-1363		-108770		-132908		-132908	
Phase		L		L		G		L		L	
Press.	Bara	1.0		1.0		0.095		0.095		1.0/1.8	
Temp	oC	45		45		45		37		37	

STREAM Nr. :		139		140	OUT	125	OUT
Name :		Recycle to R101		Purge stream		Crystal from S102	
COMP	MW	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr
Glucose	180.160	0	0.0000	0	0.0000	1.44	0.0080
Tyrosine	181.190	0	0.0000	0	0.0000	0.000	0.0000
L-Phe	165.190	0	0.0000	0	0.0000	138.576	0.8389
Acetate	59.000	0	0.0000	0	0.0000	0.638	0.0108
Biomass	24.600	0	0.0000	0	0.0000	0	0.0000
NH4	18.000	0	0.0000	0	0.0000	0.015	0.0008
H2O	18.000	34800	1933.3333	965.431	53.6351	84.331	4.6851
O2	32.000	0.00	0.0000	0.00	0.0000	0.00	0.0000
CO2	44.000	0.00	0.0000	0.00	0.0000	0.00	0.0000
N2	28.000	0.00	0.0000	0.00	0.0000	0.00	0.0000
Total		34800.00	1933.3333	965.43	53.6351	225.00	5.5436
Enthalpy	kW	-129340		-3588		-389	
Phase		L		L		S	
Press.	Bara	1.10		1.10		1.00	
Temp	oC	37		37		37	

Appendix 5.3.2 Process Stream Summary for the Batch Cycle

Stream Nr. :		101/102		IN	103/104		IN	101/102		IN	103/104		IN
Batch Cycle :		Feed Glucose to R101(1)		Start/End	Feed tyrosine to R101 (1)		Start/End	Feed Glucose to R101 (2)		Start/End	Feed tyrosine to R 101 (2)		Start/End
Component	Mw kg	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.
Glucose	0.180	1200.0	7000.000	2.936	0.0	0.000	0.000	336378.1	965.771	823.133	0.0	0.000	0.000
Tyrosine	0.181	0.0	0.000	0.000	24.0	375.000	0.059	0.0	0.000	0.000	3042.4	8.801	7.445
L-Phe	0.165	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
Acetate	0.059	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
Biomass	0.025	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
NH4	0.018	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
H2O	0.018	1014.9	5920.000	2.483	960.0	15000.000	2.349	284479.8	816.767	696.135	121696.6	352.029	297.797
O2	0.032	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
CO2	0.044	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
N2	0.028	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
Total		2214.9	7000.000	5.420	984.0	15375.000	2.408	620857.9	1782.538	1519.268	124739.0	360.830	305.242
Enthalpy :	MJ-MW-MW	-17832	-28.894	-0.012	-12959	-56.247	-0.009	-4998527	-3.986	-3.398	-1642812	-1.320037	-1.116677
Phase :	L/V/S		L			L			L			L	
Press. :	Bara		1.0/2			1.0/1.8			1.0/2			1.0/1.8	
Temp :	oC		25			25			25			25	
Cycle times	[h]												
- Cycle & Process :		-	0.2	408.7	-	0.1	408.7	-	348.3	408.7	-	345.7	408.7

(1) Filling up fermentor before fermentation

(2) Feeding fermentor during the extend growth phase and production phase

Stream Nr. :		105/106		IN	107/108		IN	109/110		IN	111		
Batch Cycle :		Feeding ammonia to R 101		Start/End	Feeding water to R101		Start/End	Mixing gas to R101		Start/End	off gas of R101		Start/End
Component	Mw kg	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.
Glucose	0.180	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
Tyrosine	0.181	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
L-Phe	0.165	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
Acetate	0.059	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
Biomass	0.025	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
NH4	0.018	6182.0	17.302	15.128	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
H2O	0.018	18546.1	51.906	45.383	70025.1	100000.000	171.355	0.0	0.000	0.000	47804.6	133.794	116.980
O2	0.032	0.0	0.000	0.000	0.0	0.000	0.000	615103.5	1721.532	1505.186	440687.5	1233.382	1078.382
CO2	0.044	0.0	0.000	0.000	0.0	0.000	0.000	487.6	1.365	1.193	254979.9	713.630	623.947
N2	0.028	0.0	0.000	0.000	0.0	0.000	0.000	681765.2	1908.103	1668.311	681765.2	1908.103	1668.311
Total		24728.1	69.208	60.511	70025.1	100000.000	171.355	1,297,356.3	3631.000	3174.690	1425237.186	3988.909	3487.620
Enthalpy :	MJ-MW-MW	-264838	-0.206	-0.180	-940438	-373.056	-0.639	-6382	-0.005	-0.004	-4699007	-3.653	-3.194
Phase :	L/V/S		L			L			V			L/V	
Press. :	Bara		1.0/1.8			1.0/1.8			1.0/2.5			1.0	
Temp :	oC		25			25			25			25	
Cycle times	[h]												
- Cycle & Process :		-	357.3	408.7	-	0.7	408.7	-	357.3	408.7		357.3	408.7

Stream Nr.	:	112			113		OUT	114/115			116		
Batch Cycle	:	Condensed water to R101		Start/End	off gas to environment		Start/End	Stream leaving R101		Start/End	Biomass recycle to R101		Start/End
Component	Mw kg	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.
Glucose	0.180	0.0	0.000	0.000	0.0	0.000	0.000	105889.8	313.422	259.117	17551.6	52.237	42.950
Tyrosine	0.181	0.0	0.000	0.000	0.0	0.000	0.000	0.1	0.000	0.000	0.0	0.000	0.000
L-Phe	0.165	0.0	0.000	0.000	0.0	0.000	0.000	317669.3	940.266	777.351	52654.9	156.711	128.849
Acetate	0.059	0.0	0.000	0.000	0.0	0.000	0.000	46950.1	138.967	114.889	7782.1	23.161	19.043
Biomass	0.025	0.0	0.000	0.000	0.0	0.000	0.000	635496.7	1881.000	1555.089	632016.0	1881.000	1546.572
NH4	0.018	0.0	0.000	0.000	0.0	0.000	0.000	1080.1	3.197	2.643	179.1	0.533	0.438
H2O	0.018	48939.4	136.970	119.757	0.0	0.000	0.000	20071695.9	59410.000	49116.357	2800224.0	8334.000	6852.276
O2	0.032	0.0	0.000	0.000	440687.5	1233.382	1078.382	58.1	0.172	0.142	9.7	0.029	0.024
CO2	0.044	0.0	0.000	0.000	254979.9	713.630	623.947	9.5	0.028	0.023	1.5	0.005	0.004
N2	0.028	0.0	0.000	0.000	681765.2	1908.103	1668.311	0.0	0.000	0.000	0.0	0.000	0.000
Total	0	48939.4	136.970	119.757	1377432.6	3855.115	3370.640	21,178,849.5	62687.052	51825.612	3,510,419.0	10447.676	8590.156
Enthalpy :	MJ-MW-MW	-657256	-0.511	-0.447	-4541395	-3.531	-3.087	-272571794	-224.106	-185.277	-40088985	-33.142	-27.250
Phase :	L/V/S		L			V			L/S			L/S	
Press. :	Bara		1.0			1.0			1.0/1.2			1.0	
Temp :	oC		25			25			37			37	
Cycle times	[h]												
- Cycle & Process :		-	357.3	408.7	-	357.3	408.7	-	337.9	408.7		336.0	408.7

Stream Nr. :		117/118/119			120 = 132+119		Calc	121			122		
Batch Cycle :		Stream leaving S101		Start/End	Stream to V101		Start/End	Liquid stream leaving V101		Start/End	Stream to V102		Start/End
Component	Mw kg	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.
Glucose	0.180	88241.5	261.185	215.931	162640.5	481.398	397.989	162640.5	481.398	397.989	162640.5	481.398	397.989
Tyrosine	0.181	0.1	0.000	0.000	0.1	0.000	0.000	0.1	0.000	0.000	0.1	0.000	0.000
L-Phe	0.165	264724.4	783.555	647.793	449466.8	1330.372	1099.866	449341.1	1330.000	1099.558	449341.1	1330.000	1099.558
Acetate	0.059	39125.1	115.806	95.741	72112.8	213.446	176.463	72112.8	213.446	176.463	72112.8	213.446	176.463
Biomass	0.025	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
NH4	0.018	900.0	2.664	2.202	1658.8	4.910	4.059	1658.8	4.910	4.059	1658.8	4.910	4.059
H2O	0.018	17254023.1	51070.000	42221.382	21612294.0	63970.000	52886.270	9530761.5	28210.000	23322.209	9530761.5	28210.000	23322.209
O2	0.032	48.3	0.143	0.118	57.8	0.171	0.141	20.3	0.060	0.050	20.3	0.060	0.050
CO2	0.044	7.8	0.023	0.019	9.2	0.027	0.022	3.1	0.009	0.007	3.1	0.009	0.007
N2	0.028	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
Total	0	17,647,070.3	52233.376	43183.187	22,298,240.1	66000.324	54564.812	10,216,538.3	30239.823	25000.336	10,216,538.3	30239.823	25000.336
Enthalpy :	MJ-MW-MW	-232235445	-190.942	-157.859	-291444837	-239.623	-198.105	-129545706	-106.511	-88.057	-129545706	-106.511	-88.057
Phase :	L/V/S		L			L			L			L	
Press. :	Bara		1.0/1.0/1.1			0.095			1.0			1.2	
Temp :	oC		37			38			45			45	
Cycle times	[h]												
- Cycle & Process :		-	337.9	408.7	-	337.9	408.7		337.9	408.7		337.9	408.7

Stream Nr. :		123			124			125			126		
Batch Cycle :		Stream leaving V102		Start/End	Stream to S102 (A/B)		Start/End	Discharge S102 (A/B)		Start/End	Feeding D101		Start/End
Component	Mw kg	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.
Glucose	0.180	162712.8	481.612	398.166	162712.8	481.612	398.166	176.3	83.938	0.431	176.3	0.522	0.000
Tyrosine	0.181	0.1	0.000	0.000	0.1	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
L-Phe	0.165	449679.0	1331.000	1100.385	449679.0	1331.000	1100.385	46047.5	21927.404	112.680	46047.5	136.296	0.000
Acetate	0.059	72144.6	213.540	176.541	72144.6	213.540	176.541	78.1	37.189	0.191	78.1	0.231	0.000
Biomass	0.025	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
NH4	0.018	1659.5	4.912	4.061	1659.5	4.912	4.061	1.8	0.874	0.004	1.8	0.005	0.000
H2O	0.018	9527383.0	28200.000	23313.941	9527383.0	28200.000	23313.941	28491.3	13567.270	69.719	28491.3	84.331	0.000
O2	0.032	20.3	0.060	0.050	20.3	0.060	0.050	0.0	0.011	0.000	0.0	0.000	35755.030
CO2	0.044	3.1	0.009	0.007	3.1	0.009	0.007	0.0	0.002	0.000	0.0	0.000	85.533
N2	0.028	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	119585.712
Total		10,213,602.4	30231.133	24993.151	10,213,602.4	30231.133	24993.151	74,795.0	35616.688	183.027	74,795.0	221.385	155426.276
Enthalpy :	MJ-MW-MW	-129508478	-106.481	-88.031	-129508478	-106.481	-88.031	-144130	-19.065	-0.098	-144130	-0.119	-0.098
Phase :	L/V/S		L/S			L/S			S			S	
Press. :	Bara		1.0			4.3			1.0			1.0	
Temp :	oC		45			45			25			25	
Cycle times	[h]												
- Cycle & Process :		-	337.9	408.7	-	337.9	408.7	-	2.1	408.7	-	337.9	408.7

Stream Nr. :		127		IN	128		OUT	129		OUT	130/131		
Batch Cycle :		Hot air to D101		Start/End	L-Phe dry product		Start/End	Off gas off D101		Start/End	Permeate leaving S102 (A/B)		Start/End
Component	Mw kg	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.
Glucose	0.180	0.0	0.000	0.000	187.1	0.554	0.458	0.0	0.000	0.000	162226.3	480.172	396.975
Tyrosine	0.181	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.1	0.000	0.000
L-Phe	0.165	0.0	0.000	0.000	46074.5	136.375	112.746	0.0	0.000	0.000	402717.8	1192.000	985.469
Acetate	0.059	0.0	0.000	0.000	82.9	0.245	0.203	0.0	0.000	0.000	71929.0	212.902	176.014
Biomass	0.025	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
NH4	0.018	0.0	0.000	0.000	1.9	0.006	0.005	0.0	0.000	0.000	1654.8	4.898	4.049
H2O	0.018	0.0	0.000	0.000	0.0	0.000	0.000	28491.3	84.331	69.719	9500355.0	28120.000	23247.802
O2	0.032	14611508.9	43248.450	35755.030	0.0	0.000	0.000	14611508.9	43248.450	35755.030	20.3	0.060	0.050
CO2	0.044	34953.8	103.459	85.533	0.0	0.000	0.000	34953.8	103.459	85.533	3.1	0.009	0.007
N2	0.028	48869424.1	144648.090	119585.712	0.0	0.000	0.000	48869424.1	144648.090	119585.712	0.0	0.000	0.000
Total		63,515,886.8	188000.000	155426.276	46,346.5	137.180	113.412	63,544,378.1	188084.331	155495.995	10,138,906.3	30010.041	24810.367
Enthalpy :	MJ-MW-MW	-	-	-	-	-	-	-	-	-	-129068278	-106.119	-87.732
Phase :	L/V/S		V			S			V			L	
Press. :	Bara		1.0			1.0			1.0			1.0/1.8	
Temp :	oC		125			25			<125			45	
Cycle times	[h]												
- Cycle & Process :		-	337.9	408.7	-	337.9	408.7	-	337.9	408.7	-	337.9	408.7

Stream Nr. :		132			133			134		OUT	135		
Batch Cycle :		Permeate recycle to V101		Start/End	Permeate recycle to R101		Start/End	Purge stream		Start/End	Vapour leaving V101		Start/End
Component	Mw kg	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.
Glucose	0.180	74399.1	220.213	182.058	85743.4	253.791	209.818	2083.5	6.167	5.098	0.0	0.000	0.000
Tyrosine	0.181	0.0	0.000	0.000	0.1	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
L-Phe	0.165	184742.4	546.817	452.073	212911.7	630.195	521.005	5173.8	15.314	12.661	0.0	0.000	0.000
Acetate	0.059	32987.7	97.640	80.722	38017.6	112.528	93.031	924.0	2.735	2.261	0.0	0.000	0.000
Biomass	0.025	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
NH4	0.018	758.8	2.246	1.857	874.7	2.589	2.140	21.3	0.063	0.052	0.0	0.000	0.000
H2O	0.018	4358271.0	12900.000	10664.888	5020457.9	14860.000	12285.290	122024.5	361.179	298.600	12081532.5	35760.000	29564.062
O2	0.032	9.5	0.028	0.023	10.8	0.032	0.026	0.3	0.001	0.001	0.0	0.000	0.000
CO2	0.044	1.4	0.004	0.003	1.6	0.005	0.004	0.0	0.000	0.000	0.0	0.000	0.000
N2	0.028	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
Total		4,651,169.8	13766.948	11381.625	5,358,017.8	15859.140	13111.314	130,227.5	385.459	318.673	12,081,532.5	35760.000	29564.062
Enthalpy :	MJ-MW-MW	-59209392	-48.681	-40.247	-68207566	-56.080	-46.363	-1657796	-1.363	-1.127	-132292781	-108.770	-89.924
Phase :	L/V/S		L			L			L			G	
Press. :	Bara		1.1			1.0			1.0			0.095	
Temp :	oC		45			45			45			45	
Cycle times	[h]												
- Cycle & Process :		-	337.9	408.7	-	337.9	408.7	-	337.9	408.7	-	337.9	408.7

Stream Nr.	:	136			137/138			139			140		OUT
Batch Cycle	:	After E102		Start/End	Stream leaving V103		Start/End	Recycle to R101		Start/End	Purge stream		Start/End
Component	Mw kg	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.
Glucose	0.180	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
Tyrosine	0.181	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
L-Phe	0.165	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
Acetate	0.059	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
Biomass	0.025	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
NH4	0.018	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
H2O	0.018	12081532.5	35760.000	29564.062	12081532.5	35760.000	29564.062	11757196.1	34800.000	28770.396	326171.3	965.431	798.156
O2	0.032	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
CO2	0.044	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
N2	0.028	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
Total		12,081,532.5	35760.000	29564.062	12,081,532.5	35760.000	29564.062	11,757,196.1	34800.000	28770.396	326,171.3	965.431	798.156
Enthalpy :	MJ-MW-MW	-161650905	-132.908	-109.880	-161650905	-132.908	-109.880	-157311284	-129.340	-106.930	-4364172	-3.588	-2.966
Phase :	L/V/S		L			L			L			L	
Press. :	Bara		0.095			1.0/1.8			1.10			1.10	
Temp :	oC		37			37			37			37	
Cycle times	[h]												
- Cycle & Process :		-	337.9	408.7	-	337.9	408.7	-	337.9	408.7	-	337.9	408.7

Stream Nr. :		141		IN	142/143		OUT	144/145		IN	146		OUT
Batch Cycle :		Air into V103		Start/End	Air leaving V103		Start/End	Washing water to S102 (A/B)		Start/End	Washing water leaving S102 (A/B)		Start/End
Component	Mw kg	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.
Glucose	0.180	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	310.2	147.731	0.759
Tyrosine	0.181	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
L-Phe	0.165	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	770.4	366.865	1.885
Acetate	0.059	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	137.5	65.453	0.336
Biomass	0.025	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
NH4	0.018	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.005	0.000
H2O	0.018	0.0	0.000	0.000	0.0	0.000	0.000	30240.0	14400.000	73.999	30240.0	14400.000	73.999
O2	0.032	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
CO2	0.044	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
N2	0.028	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000	0.0	0.000	0.000
Total		0.0	0.000	0.000	0.0	0.000	0.000	30,240.0	14400.000	73.999	31,458.1	14980.054	76.979
Enthalpy :	MJ-MW-MW	-	-	-	-	-	-	-406123	-53.720	-0.276	-410528	-54.303	-0.279
Phase :	L/V/S		G			G			L			L	
Press. :	Bara		1.0			0.055/1.0			1.0/4.3			1.00	
Temp :	oC		25			25			25			25	
Cycle times	[h]												
- Cycle & Process :		-	337.9	408.7	-	337.9	408.7	-	2.1	408.7	-	2.1	408.7

Stream Nr. :		147		IN	148		IN
Batch Cycle :		Crystal seed to V102		Start/End	Inoculum to R101		Start/End
Component	Mw kg	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.
Glucose	0.180	0.0	0.000	0.000	0.0	0.000	0.000
Tyrosine	0.181	0.0	0.000	0.000	0.0	0.000	0.000
L-Phe	0.165	0.8	0.000	0.002	0.0	0.000	0.000
Acetate	0.059	0.0	0.000	0.000	0.0	0.000	0.000
Biomass	0.025	0.0	0.000	0.000	0.0	0.000	0.000
NH4	0.018	0.0	0.000	0.000	0.0	0.000	0.000
H2O	0.018	0.0	0.000	0.000	8000.0	27000.000	66.070
O2	0.032	0.0	0.000	0.000	0.0	0.000	0.000
CO2	0.044	0.0	0.000	0.000	0.0	0.000	0.000
N2	0.028	0.0	0.000	0.000	0.0	0.000	0.000
Total		0.8	0.000	0.002	8,000.0	27000.000	66.070
Enthalpy :	MJ-MW-MW	0	0.000	0.000	-107440	-100.725	-0.073
Phase :	L/V/S		S			L/S	
Press. :	Bara		1.0			1.00	
Temp :	oC		25			25	
Cycle times	[h]						
- Cycle & Process :		-	0.0	408.7	-	0.3	408.7

Appendix 5.4 Utilities ISPR

SUMMARY OF UTILITIES				
EQIPMENT	**UTILITIES**			
Unit	**Power**	**HEATING**	**COOLING**	
	Electricity [MJ]	Heating (T =) [ton]	Cooling water (T=10) [ton]	
		Pumps		
Pumps				
P101	27			
P102	27			
P103	3.5			
P104	2990			
P105	608			
P106	113			
P107	134			
P108	4040			
P109	3300			
P100	125			
P111	13500			
		Condensers		
E102			7717248	
		Fermentor		
R101: stirring	6.58E+6			
R101: cooling			678142.1	
		Evaporator		
V101		282130.9		
		Compressor		
K101	1.29E+6			
		Centrifuge		
S101	1.20E+5			
		Crystallizer		
			239.7	
Total	7996800.3	282130.9	8395629.7	
Price (€/Unit)	0.022	11.7	0.010	
Cost (k€)	176.9	3305.9	82.6	
Total cost (k€)				**3565.3**

Project ID number: CPD 3324
Complete date: 16-10-2005

APPENDIX 6 SPECIFICATION SHEETS

PUMPS, BLOWERS & COMPRESSORS – SUMMARY

EQUIPMENT NR. : NAME **:**	**K-101(A/B)** **Air compressor:**				
Type: **Number:**	Centrifugal 2				
Medium Transferred:	Air				
Capacity **[kg/s] :** **[m^3/s] :**	 0.542				
Density [kg/m^3]:	1.15				
Pressure [bara] **Suct. / Disch. :** **Diff. Pressure/ stage [bar] :** **Stages :** **Speed [rpm] :**	 1.01 / 2.533 0.7 4 (2) 2550				
Temperature **In / Out [°C]:**	 25 / 25				
Power [kW] **-Theor.:** **-Actual:**	 61.34 n.a.				
Number **-Theor.:** **-Actual:**	 2 *(1)*				
Special Materials of Construction:	MS casing				
Other:					

Remarks:

(1) One installed spare included.

(2) Based on table 10.16 Coulson & Richardson's

Designers :	M.H. Ucisik H.Yang J.Nie J.F.Solarte	Project ID-Number : Date :	**CPD3324** October 11 2005

Pump Summary

EQUIPMENT NR. : **NAME :**	**P101** **Feedng Glucose Pump**	**P102** **Feeding Tyrosine Pump**	**P103** **Feeding Ammonia Water Pump**	**P104** **Filling R101 Pump**
Type : **Number :**	Peristaltic 2	Centrifugal 2	Diaphram 2	Rotary (sliding vane) 2
Capacity [m^3/s] :	2.966E-4	9.664E-5	2.01E-5	7.64E-4
Density [kg/m^3] :	1290	1000	1000	1000
Pressure [bara] Suct./ Disch. :	1.0/2.0	1.0/1.8	1.0/1.8	1.0/1.8
Temperature In/Out [^{o}C] :	25/25	25/25	25/25	25/25
Power [kW] **-Theor. :** **-Actual :**	 1.46 2.17	 0.014 0.021	 0.0018 0.0027	 1.67 2.12
Special Materials of Construction :	MS casing	MS casing	MS casing	MS casing

EQUIPMENT NR.: **NAME :**	**P105** **R101 Withdraw Pump**	**P106** **V105 Withdraw Pump**	**P107** **V101 Withdraw Pump**	**P108** **V102 Withdraw Pump**
Type : **Number :**	Rotary (sliding vane) 2	Centrifugal 2	Centrifugal 2	Centrifugal 2
Capacity [m^3/s] :	0.017	0.015	8.4E-3	8.398E-3
Density [kg/m^3] :	1000	1000	1000	1000
Pressure [bara] Suct./ Disch. :	1.0/1.2	1.0/1.1	1.0/1.2	1.0/4.5
Temperature In/Out [^{o}C] :	37/37	37/37	45/45	45/45
Power [kW] **-Theor. :** **-Actual :**	 0.34 0.47	 0.06 0.089	 0.73 1.04	 2.04 3.13
Special Materials of Construction :	MS casing	MS casing	MS casing	MS casing

EQUIPEMENT NR.: **NAME :**	**P109** **Effluent Withdraw Pump**	**P110** **V103 Withdraw Pump**	**P111** **Feeding Wash Water Pump**
Type : **Number :**	Centrifugal 2	Centrifugal	Centrifugal
Capacity [m^3/s] :	8.336E-3	9.933E-3	0.028
Density [kg/m^3] :	1000	1000	1000
Pressure [bara] Suct./ Disch. :	1.0/1.8	1.0/1.8	1.0/4.5
Temperature In/Out [°C] :	45/45	25/25	25/25
Power [kW]			
-Theor. :	1.5	0.066	7.1
-Actual :	2.6	0.098	10.1
Special Materials of Construction :	MS casing	MS casing	MS casing

CENTRIFUGE – SPECIFICATION SHEET

EQUIPMENT NUMBER : S-101
Operating : 1
NAME : Disk Centrifuge
Installed Spare : 0

Service : Biomass separation
Type : Disk
Number : 1

Operating Conditions & Physical Data

Pumped liquid : Fermentation broth
Temperature (T) **[^oC]** : 37.0
Density (ρ) **[kg/m^3]** : 1000
Viscosity(η) **[N·s/m^2]** : 8.937E-4

Power (1)

Max. Capacity **(Φ_v)** **[m^3/s]** : 0.025
Theoretical Power **[kW]** : 93.212

Construction Details *(1)*

RPM	: 4200	**Nominal diameter [m] :**	0.495
Drive	: Electrical	**Number of disk :**	144
Type electrical motor	: **Yes**	**Distance between disk [m] :**	0.002
Rotational direction	: **Clock /** **~~Counter Cl.~~**	**Discharge Nozzle [-] :**	yes
		Disc angle [-] :	**n.a.**
Flexible Coupling	: n.a.		
Pressure Gauge Discharge	: n.a.		

Construction Materials *(2)*

Centrifuge House	:SS	**Wear Rings :**	**n.a.**
Rotor	:SS	**Shaft Box :**	**n.a.**
Special provisions	:none		
Operating Pressure [bara]	: 1	**Test Pressure [bara] :**	**n.a.**

Remarks:
(1) Perry's Chemical Engineers' Handbook
(2) SS = Stainless steel; HT Steel = High Tensile Steel

Designers :	M.H. Ucisik H.Yang J.Nie J.F.Solarte	Project ID-Number :	**CPD3324**
		Date :	October 11 2005

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER :	**V-103**	**In Series :**	1
NAME :	**E-102 Condenser**	**In Parallel :**	none

General Data

Service		:	**Condenser (Water cooled)** *(1), (2)*	
Type		:	**Floating Head**	
Position		:	**Horizontal**	
Capacity	**[kW]**	:	24120	**(Calc.)**
Heat Exchange Area	**[m^2]**	:	2452 *(3)*	**(Calc.)**
Overall Heat Transfer Coefficient	**[$W/m^2 \cdot {}^oC$]**	:	440.816	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^oC]**	:	25.62	
Passes Tube Side		:	4	
Passes Shell Side		:	1	
Correction Factor LMTD (min. 0.75)		:	0.96	
Corrected LMTD	[^oC]	:	24.595	

Process Conditions

			Shell Side	**Tube Side**
Medium		:	Water Steam	Water
Mass Stream	**[kg/s]**	:	9.933	288.343
Mass Stream to - Condense	**[kg/s]**	:	9.933.	n.a
Average Specific Heat	**[$kJ/kg \cdot {}^oC$]**	:	4.181	4.19
Heat of Condensation	**[kJ/kg]**	:	2442	-
Temperature IN	**[^oC]**	:	45	5
Temperature OUT	**[^oC]**	:	37	25
Pressure	**[bara]**	:	0.095	1.013
Material		:	CS	CS *(4)*

Remarks:

(1) Calculation of Condenser: "Coulson and Richardson's Chemical Engineering", R.K.Sinnott
(2) Equipment Size: Inside Shell Diameter: 3.513 m ; Tube Length: 1.83 m ;
(3) Tube Sizes: inside: 18 mm ; outside 20mm ; Square pitch
(4) CS: Carbon steel

Designers : J. Nie J.F.S.Vasquez M.H.Ucisik H.Huiling	Project ID-Number : **CPD3324** Date : October 11th 2005

HEAT EXCHANGER – SPECIFICATION SHEET

EQUIPMENT NUMBER :	**V-102**	**In Series :**	1
NAME :	**V- 101 Evaporator**	**In Parallel :**	none

General Data

Service	: *(1), (2)*	**Vaporizer (Water evaporated)**
Type	:	**Floating Head**
Position	:	**Vertical**

Capacity	**[kW]**	:	24270	**(Calc.)**
Heat Exchange Area	**[m^2]**	:	375.895	*(3)* **(Calc.)**
Overall Heat Transfer Coefficient	**[$W/m^2 \cdot {}^oC$]**	:	3589	**(Approx.)**
Log. Mean Temperature Diff. (LMTD)	**[^oC]**	:	58.161	
Passes Tube Side		:	8	
Passes Shell Side		:	1	
Correction Factor LMTD (min. 0.75)		:	1	
Corrected LMTD	**[^oC]**	:	58.108	

Process Conditions

			Shell Side	**Tube Side**
Medium		:	Process Liquid	Water
Mass Stream	**[kg/s]**	:	18.336	10.753
Mass Stream to				
- Condense	**[kg/s]**	:	-.	10.753
- Evaporate	**[kg/s]**	:	9.933	n.a
Average Specific Heat	**[$kJ/kg \cdot {}^oC$]**	:	4.18	3.054
Heat of Vaporization	**[kJ/kg]**	:	2295	-
Condensation	**[kJ/kg]**	:	-	2257
Temperature IN	**[^oC]**	:	38.67	100
Temperature OUT	**[^oC]**	:	45	100
Pressure	**[bara]**	:	0.095	1.013
Material		:	CS	CS

Remarks:

(1) Calculation of Condenser: "Coulson and Richardson's Chemical Engineering", R.K.Sinnott

(2) Equipment Size: Inside Shell Diameter: 0.861 m ; Tube Length: 8 m ;

(3) Tube Diameter: inside: 14.83 mm ; outside 19.05mm ; Triangular square patterns

Designers : J. Nie J.F.S.Vasquez M.H.Ucisik H.Huiling	Project ID-Number : **CPD3324** Date : October 11th 2005

CRYSTALLIZER – SPECIFICATION SHEET

EQUIPMENT NR. : **V-102 :**	**V-102** **Crystallizer**

General Data

Service :	Crystallization
Vessel Diameter [m]:	1.017
Vessel Hight [m]:	1.526
Total Volume [m^3]:	1.241
Position :	Vertical
Heating/Cooling :	Jacket
Heating/Cooling medium:	Cooling Water (1)

Process conditions

Operational Details	Crystallization
Temp. [°C] :	45 / 45
Pressure [bar] :	1
Density [kg/ m^3]:	1000
Heating/Cooling	Cooling
Batch/Continuous	Continuous
Number - Series - **Parallel:**	1 -
Materials of Construction:	SS (2)

Other :					

Remarks:

(1) This specification sheet describes the propertyes of crystallization vessel unit of the integrated forced circulation evaporation crystallizer. The cooling duty mentioned is only for the removal of the heat of crystallization and when it is compared to the duty of evaporation unit it is negligible. This separation of the evaporation crystallizer is done for detailed calculations of the total equipment.

(2) SS = Stainless Steel.

Designers : J. Nie J.F.S.Vasquez M.H.Ucisik H.Huiling	Project ID-Number : **CPD3324** Date : October 11th 2005

CENTRIFUGAL PUMP – SPECIFICATION SHEET

EQUIPMENT NUMBER :	**P101**	**Operating**	**:** 1
NAME	**: Feeding Glucose Pump**	**Installed Spare**	**:** 1

Service : Feeding glucose pump
Type : peristaltic pump
Number : 2

Operating Conditions & Physical Data

Pumped liquid			:	Glucose
Temperature	(T)	**[°C]**	:	25
Density	(ρ)	**[kg/m^3]**	:	1290
Viscosity	(η)	**[N·s/m^2]**	:	8
Vapour Pressure	(p_v)	**[bara]**	:	0.032
	at Temperature [°C]		:	25

Capacity	(Φ_v)	**[m^3/s]**	:	2.966E-4
Suction Pressure	(p_s)	**[bara]**	:	1
Discharge Pressure	(p_d)	**[bara]**	:	2.0
Theoretical Power		**[kW]**	:	1.46
Pump Efficiency		**[-]**	:	0.67
Power at Shaft		**[kW]**	:	2.17

Construction Details *(1)*

RPM	:900	**Nominal diameter**	
Drive	:Electrical	**Suction Nozzle** [...]	:
Type electrical motor	:	**Discharge Nozzle** [...]	:
Tension [V]	:380	**Cooled Bearings**	**:Yes/No**
Rotational direction	**:Clock /**	**Cooled Stuffing Box**	**:Yes/No**
		Smothering Gland	**:Yes/No**
Foundation Plate	**:Combined**	**If yes**	
Flexible Coupling	:Yes	**- Seal Liquid**	**:Yes/No**
Pressure Gauge Suction	:No	**- Splash Rings**	**:Yes/No**
Pressure Gauge Discharge	:Yes	**- Packing Type**	**: - Mechanical**
Min. Overpressure above		**Seal**	**:Yes/No**
p_v/p_m **[bar]**	: 0.1	**- N.P.S.H. [m]**	**:{ =** $p_m \cdot \rho g$ **}**

Construction Materials *(2)*

Pump House	: MS	**Wear Rings**	:
Pump Rotor	: HT Steel	**Shaft Box**	:
Shaft	: HT Steel		
Special provisions	: none		
Operating Pressure [bara]	: 1	**Test Pressure [bara]**	:

Remarks:

(3) Double mechanical seals and seal fluid required for LPG service. Further details to be specified by Rotating Equipment specialist.

(4) MS = Mild Steel; HT Steel = High Tensile Steel

CENTRIFUGAL PUMP – SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**P102**	**Operating** :	1
NAME	:	**Feeding Tyrosine Pump**	**Installed Spare** :	1

Service : Feeding tyrosine pump
Type : Centrifugal **Number** : 2

Operating Conditions & Physical Data

Pumped liquid			:	Tyrosine
Temperature	(T)	**[°C]**	:	25
Density	(ρ)	**[kg/m^3]**	:	1000
Viscosity	(η)	**[N·s/m^2]**	:	7.64E-4
Vapour Pressure	(p_v)	**[bara]**	:	0.032
		at Temperature [°C]	:	25

Power

Capacity	(Φ_v)	**[m^3/s]**	:	9.664E-5
Suction Pressure	(p_s)	**[bara]**	:	1
Discharge Pressure	(p_d)	**[bara]**	:	1.8
Theoretical Power		**[kW]**	:	0.014
Pump Efficiency		**[-]**	:	0.7
Power at Shaft		**[kW]**	:	0.021

Construction Details *(1)*

RPM		:	900	**Nominal diameter**			
Drive		:	Electrical	**Suction Nozzle**	[…]	:	
Type electrical motor		:		**Discharge Nozzle**	[…]	:	
Tension	[V]	:	380	**Cooled Bearings**		:	**Yes / No**
Rotational direction		:	**Clock**	**Cooled Stuffing Box**		:	**Yes / No**
Foundation Plate		:	**Combined**	**Smothering Gland**		:	**Yes / No**
				If yes			
Flexible Coupling		:	Yes	**- Seal Liquid**		:	**Yes / No**
Pressure Gauge Suction		:	No	**- Splash Rings**		:	**Yes / No**
Pressure Gauge Discharge		:	Yes	**- Packing Type**		:	
Min. Overpressure above				**- Mechanical Seal**		:	**Yes / No**
p_v/p_m	**[bar]**	:	0.1	**- N.P.S.H.**	**[m]**	:	**8.2**

Construction Materials *(2)*

Pump House		:	MS	**Wear Rings**		:
Pump Rotor		:	HT Steel	**Shaft Box**		:
Shaft		:	HT Steel			
Special provisions		:	none			
Operating Pressure	**[bara]**	:	1	**Test Pressure**	**[bara]**	:

Remarks: (5) Double mechanical seals and seal fluid required for LPG service.Further details to be specified by Rotating Equipment specialist. MS = Mild Steel; T Steel = High Tensile Steel

Designers :	Juan Felipe Solarte	Huiling Yang	Project ID-Number :	**CPD3324**
	Mehmet Hikmet Ucisik	Jing Nie	Date :	October 15th 2005

CENTRIFUGAL PUMP – SPECIFICATION SHEET

EQUIPMENT NUMBER	**: P103**	**Operating**	**:**	1
NAME	**: Feeding Ammonia Water Pump**	**Installed Spare**	**:**	1

Service	: Feeding ammonia water pump			
Type	: diaphragm	**Number**	**:**	2

Operating Conditions & Physical Data

Pumped liquid			:	ammonia water
Temperature	(T)	**[°C]**	:	25
Density	(ρ)	**[kg/m^3]**	:	1000
Viscosity	(η)	**[N·s/m^2]**	:	7.64E-4
Vapour Pressure	(p_v)	**[bara]**	:	0.032
	at Temperature	**[°C]**	:	25

Power

Capacity	(Φ_v)	**[m^3/s]**	:	2.01E-5
Suction Pressure	(p_s)	**[bara]**	:	1
Discharge Pressure	(p_d)	**[bara]**	:	1.8
Theoretical Power		**[kW]**	:	0.0018
Pump Efficiency		**[-]**	:	0.67
Power at Shaft		**[kW]**	:	0.0027

Construction Details *(1)*

RPM		:	900	**Nominal diameter**			
Drive		:	Electrical	**Suction Nozzle**	**[…]**	**:**	
Type electrical motor		:		**Discharge Nozzle**	**[…]**	**:**	
Tension	**[V]**	:	380	**Cooled Bearings**		**:**	**Yes / No**
Rotational direction		:	**Clock /**	**Cooled Stuffing Box**		**:**	**Yes / No**
				Smothering Gland		**:**	**Yes / No**
~~**Counter Cl.**~~				**If yes**			
Foundation Plate		:	**Combined**	**- Seal Liquid**		**:**	**Yes / No**
/			~~**two parts**~~	**- Splash Rings**		**:**	**Yes / No**
Flexible Coupling		:	Yes	**- Packing Type**		**:**	
Pressure Gauge Suction		:	No	**- Mechanical Seal**		**:**	**Yes / No**
Pressure Gauge Discharge		:	Yes	**- N.P.S.H.**	**[m]**	**:**	**10.1**
Min. Overpressure above				$\{ = p_m \cdot \rho g \}$			
p_v/p_m	**[bar]**	:	0.1				

Construction Materials *(2)*

Pump House		:	MS	**Wear Rings**		**:**	
Pump Rotor		:	HT Steel	**Shaft Box**		**:**	
Shaft		:	HT Steel				
Special provisions		:	none				
Operating Pressure	**[bara]**	:	1	**Test Pressure**	**[bara]**	**:**	

Remarks:

(5) Double mechanical seals and seal fluid required for LPG service. Further details to be specified by Rotating Equipment specialist.

(6) MS = Mild Steel; HT Steel = High Tensile Steel

CENTRIFUGAL PUMP – SPECIFICATION SHEET

EQUIPMENT NUMBER :	**P104**	**Operating**	**:**	1
NAME	**: Filling R101 Pump**	**Installed Spare**	**:**	1

Service : Filling water pump
Type : rotary (sliding vane)
Number : 2

Operating Conditions & Physical Data

Pumped liquid			:	Water
Temperature	(T)	[°C]	:	25
Density	(ρ)	[kg/m^3]	:	1000
Viscosity	(η)	[N·s/m^2]	:	7.64E-4
Vapour Pressure	(p_v)	[bara]	:	0.032
	at Temperature [°C]		:	25

Power

Capacity	(Φ_v)	[m^3/s]	:	0.019
Suction Pressure	(p_s)	[bara]	:	1
Discharge Pressure	(p_d)	[bara]	:	1.8
Theoretical Power		[kW]	:	1.67
Pump Efficiency		[-]	:	0.72
Power at Shaft		[kW]	:	2.32

Construction Details *(1)*

RPM		:	900	**Nominal diameter**			
Drive		:	Electrical	**Suction Nozzle**	[…]	:	
Type electrical motor		:		**Discharge Nozzle**	[…]	:	
Tension	[V]	:	380	**Cooled Bearings**		:	**Yes / No**
Rotational direction		:	**Clock /**	**Cooled Stuffing Box**		:	**Yes / No**
				Smothering Gland		:	**Yes / No**
~~Counter Cv el.~~				**If yes**			
Foundation Plate		:	**Combined**	**- Seal Liquid**		:	**Yes / No**
/			**~~two parts~~**	**- Splash Rings**		:	**Yes / No**
Flexible Coupling		:	Yes	**- Packing Type**		:	
Pressure Gauge Suction		:	No	**- Mechanical Seal**		:	**Yes / No**
Pressure Gauge Discharge		:	Yes	**- N.P.S.H.**	[m]	:	**9.7**
Min. Overpressure above				{ $= p_m \cdot \rho g$ }			
p_v/p_m	[bar]	:	0.1				

Construction Materials *(2)*

Pump House		:	MS	**Wear Rings**		:	
Pump Rotor		:	HT Steel	**Shaft Box**		:	
Shaft		:	HT Steel				
Special provisions		:	none				
Operating Pressure	[bara]	:	1	**Test Pressure**	[bara]	:	

Remarks:

(7) Double mechanical seals and seal fluid required for LPG service. Further details to be specified by Rotating Equipment specialist.

MS = Mild Steel; HT Steel = High Tensile Steel

CENTRIFUGAL PUMP – SPECIFICATION SHEET

EQUIPMENT NUMBER :	**P105**	**Operating**	**:1**
NAME :	**R101 Withdraw Pump**	**Installed Spare**	**:1**
Service :	**R101 Withdraw Pump**		
Type :	rotary (sliding vane) 1	**Number :**	2

Operating Conditions & Physical Data

Pumped liquid		: Mixed Liquid
Temperature (T)	**[°C]**	: 37
Density (ρ)	**[kg/m^3]**	: 1000
Viscosity (η)	**[N·s/m^2]**	: 6.276E-4
Vapour Pressure (p_v)	**[bara]**	: 0.063
at Temperature	**[°C]**	: 37

Power

Capacity	**(Φ_v)**	**[m^3/s]**	: 0.017
Suction Pressure	**(p_s)**	**[bara]**	: 1
Discharge Pressure	**(p_d)**	**[bara]**	: 1.2
Theoretical Power		**[kW]**	: 0.34
Pump Efficiency		**[-]**	: 0.72
Power at Shaft		**[kW]**	: 0.47

Construction Details *(1)*

RPM		: 900	**Nominal diameter**		
Drive		: Electrical	**Suction Nozzle**	**[…]**	**:**
Type electrical motor		:	**Discharge Nozzle**	**[…]**	**:**
Tension	**[V]**	: 380	**Cooled Bearings**		**:Yes / No**
Rotational direction		: **Clock / ~~Counter Cl.~~**	**Cooled Stuffing Box**		**:Yes / No**
			Smothering Gland		**:Yes / No**
Foundation Plate		: **Combined / ~~two parts~~**	**If yes**		
			- Seal Liquid		**:Yes / No**
Flexible Coupling		: Yes	**- Splash Rings**		**:Yes / No**
Pressure Gauge Suction		: No	**- Packing Type**		**: -**
Pressure Gauge Discharge		: Yes	**Mechanical Seal**		**:Yes / No**
Min. Overpressure above			**- N.P.S.H.**	**[m]**	**: 11.0**
p_v/p_m	**[bar]**	: 0.1	**$\{=p_m \cdot \rho g\}$**		

Construction Materials *(2)*

Pump House		: MS	**Wear Rings**		**:**
Pump Rotor		: HT Steel	**Shaft Box**		**:**
Shaft		: HT Steel			
Special provisions		: none			
Operating Pressure	**[bara]**	: 1	**Test Pressure**	**[bara]**	**:**

Remarks:

(8) Double mechanical seals and seal fluid required for LPG service. Further details to be specified by Rotating Equipment specialist.

(9) MS = Mild Steel; HT Steel = High Tensile Steel

CENTRIFUGAL PUMP – SPECIFICATION SHEET

EQUIPMENT NUMBER :	**P106**	**Operating**	**:**	1
NAME :	**V105 Withdraw Pump**	**Installed Spare**	**:**	1
Service :	**V105 Withdraw Pump**			
Type :	Centrifugal	**Number**	**:**	2

Operating Conditions & Physical Data

Pumped liquid			:	Mixed liquid
Temperature	**(T)**	**[°C]**	:	37
Density	**(ρ)**	**[kg/m^3]**	:	1000
Viscosity	**(η)**	**[N·s/m^2]**	:	6.276E-4
Vapour Pressure	**(p_v)**	**[bara]**	:	0.063
		at Temperature [°C]	:	37

Power

Capacity	**(Φ_v)**	**[m^3/s]**	:	0.015
Suction Pressure	**(p_s)**	**[bara]**	:	1
Discharge Pressure	**(p_d)**	**[bara]**	:	1.1
Theoretical Power		**[kW]**	:	0.06
Pump Efficiency		**[-]**	:	0.68
Power at Shaft		**[kW]**	:	0.089

Construction Details *(1)*

RPM		:	900	**Nominal diameter**			
Drive		:	Electrical	**Suction Nozzle**	**[…]**	**:**	
Type electrical motor		:		**Discharge Nozzle**	**[…]**	**:**	
Tension	**[V]**	:	380	**Cooled Bearings**		**:**	**Yes / No**
Rotational direction		:	**Clock**	**Cooled Stuffing Box**		**:**	**Yes / No**
				Smothering Gland		**:**	**Yes / No**
Foundation Plate		:	**Combined**	**If yes**			
Flexible Coupling		:	Yes	**- Seal Liquid**		**:**	**Yes / No**
Pressure Gauge Suction		:	No	**- Splash Rings**		**:**	**Yes / No**
Pressure Gauge Discharge		:	Yes	**- Packing Type**		**:**	
Min. Overpressure above				**- Mechanical Seal**		**:**	**Yes / No**
p_v/p_m	**[bar]**	:	0.1	**- N.P.S.H.**	**[m]**	**:**	**10.4**
				$\{=p_m \cdot \rho g\}$			

Construction Materials *(2)*

Pump House		:	MS	**Wear Rings**		**:**
Pump Rotor		:	HT Steel	**Shaft Box**		**:**
Shaft		:	HT Steel			
Special provisions		:	none			
Operating Pressure	**[bara]**	:	1	**Test Pressure**	**[bara]**	**:**

Remarks:

(10) Double mechanical seals and seal fluid required for LPG service. Further details to be specified by Rotating Equipment specialist.

MS = Mild Steel; HT Steel = High Tensile Steel

EQUIPMENT NUMBER :	**P107**	**Operating**	**:**	1
NAME	**: V101 Withdraw Pump**	**Installed Spare**	**:**	1

Service	**: V101 Withdraw Pump**			
Type	**:** Centrifugal	**Number**	**:**	2

Operating Conditions & Physical Data

Pumped liquid	:	Mixed liquid		
Temperature	(T)	**[^{o}C]**	**:**	45
Density	(ρ)	**[kg/m^3]**	**:**	1000
Viscosity	(η)	**[N·s/m^2]**	**:**	5.551E-4
Vapour Pressure	(p_v))	**[bara]**	**:**	0.096
		at Temperature [^{o}C]	**:**	45

Power

Capacity	(Φ_v)	**[m^3/s]**	**:**	8.4E-3
Suction Pressure	(p_s)	**[bara]**	**:**	1
Discharge Pressure	(p_d)	**[bara]**	**:**	1.2
Theoretical Power		**[kW]**	**:**	0.73
Pump Efficiency	**[-]**		**:**	0.7
Power at Shaft	**[kW]**		**:**	1.04

Construction Details *(1)*

RPM		**:**	900	**Nominal diameter**			
Drive		**:**	Electrical	**Suction Nozzle**	**[…]**	**:**	
Type electrical motor		**:**		**Discharge Nozzle**	**[…]**	**:**	
Tension	**[V]**	**:**	380	**Cooled Bearings**		**:**	**Yes / No**
Rotational direction		**:**	**Clock**	**Cooled Stuffing Box**		**:**	**Yes / No**
Foundation Plate		**:**	**Combined**	**Smothering Gland**		**:**	**Yes / No**
Flexible Coupling		**:**	Yes	**If yes**			
Pressure Gauge Suction		**:**	No	**- Seal Liquid**		**:**	**Yes / No**
Pressure Gauge Discharge		**:**	Yes	**- Splash Rings**		**:**	**Yes / No**
Min. Overpressure above				**- Packing Type**		**:**	
p_v/p_m	**[bar]**	**:**	0.1	**- Mechanical Seal**		**:**	**Yes / No**
				- N.P.S.H.	**[m]**	**:**	**3.2**
				$\{=p_m \cdot \rho g\}$			

Construction Materials *(2)*

Pump House		**:**	MS	**Wear Rings**		**:**
Pump Rotor		**:**	HT Steel	**Shaft Box**		**:**
Shaft		**:**	HT Steel			
Special provisions		**:**	none			
Operating Pressure	**[bara]**	**:**	1	**Test Pressure**	**[bara]**	**:**

Remarks:

(11) Double mechanical seals and seal fluid required for LPG service. Further details to be specified by Rotating Equipment specialist.

MS = Mild Steel; HT Steel = High Tensile Steel

EQUIPMENT NUMBER : **P108** **Operating :** 1
NAME : **V102 Withdraw Pump** **Installed Spare :** 1

Service : **V102 Withdraw Pump**
Type : Centrifugal
Number : 2

Operating Conditions & Physical Data

Pumped liquid : Mixed Liquid
Temperature **(T) [°C] :** 45
Density **(ρ) [kg/m^3] :** 1000
Viscosity **(η) [N·s/m^2]:** 0.555
Vapour Pressure **(p_v) [bara] :** 0.096
at Temperature [°C] : 45

Power

Capacity **(Φ_v) [m^3/s] :** 8.398E-3
Suction Pressure **(p_s) [bara] :** 1
Discharge Pressure **(p_d) [bara] :** 4.5

Theoretical Power **[kW] :** 2.04

Pump Efficiency **[-] :** 0.65
Power at Shaft **[kW] :** 3.13

Construction Details *(1)*

RPM		:	900	**Nominal diameter**			
Drive		:	Electrical	**Suction Nozzle**	**[…]**	:	
Type electrical motor		:		**Discharge Nozzle**	**[…]**	:	
Tension	**[V]**	:	380	**Cooled Bearings**		:	**Yes / No**
Rotational direction		:	**Clock / ~~Counter Cl.~~**	**Cooled Stuffing Box**		:	**Yes / No**
				Smothering Gland		:	**Yes / No**
				If yes			
Foundation Plate /		:	**Combined ~~two parts~~**	**- Seal Liquid**		:	**Yes / No**
				- Splash Rings		:	**Yes / No**
Flexible Coupling		:	Yes	**- Packing Type**		:	
Pressure Gauge Suction		:	No	**- Mechanical Seal**		:	**Yes / No**
Pressure Gauge Discharge		:	Yes	**- N.P.S.H.**	**[m]**	:	**10**
Min. Overpressure above				**{ $= p_m \cdot \rho g$ }**			
p_v/p_m	**[bar]**	:	0.1				

Construction Materials *(2)*

Pump House		:	MS	**Wear Rings**		:	
Pump Rotor		:	HT Steel	**Shaft Box**		:	
Shaft		:	HT Steel				
Special provisions		:	none				
Operating Pressure	**[bara]**	:	1	**Test Pressure**	**[bara]**	:	

Remarks:

(12) Double mechanical seals and seal fluid required for LPG service. Further details to be specified by Rotating Equipment specialist.

(13) MS = Mild Steel; HT Steel = High Tensile Steel

EQUIPMENT NUMBER :	**P109**	**Operating** :	1
NAME :	**Effluent Withdraw Pump**	**Installed Spare** :	1
Service :	**Effluent Withdraw Pump**		
Type :	Centrifugal	**Number** :	2
Operating Conditions & Physical Data			
Pumped liquid :	Mixed Liquid		
Temperature (T) **[°C]** :	45		
Density (ρ) **[kg/m³]** :	1000		
Viscosity (η) **[N·s/m²]** :	0.555		
Vapour Pressure (p_v) **[bara]** :	0.096		
at Temperature [°C] :	45		
Power			
Capacity (Φ_v) **[m³/s]** :	8.336E-3		
Suction Pressure (p_s) **[bara]** :	1		
Discharge Pressure (p_d) **[bara]** :	1.8		
Theoretical Power [kW] :	1.5		
Pump Efficiency [-] :	0.6		
Power at Shaft [kW] :	2.6		
Construction Details *(1)*			
RPM :	900	**Nominal diameter**	
Drive :	Electrical	**Suction Nozzle [...]** :	
Type electrical motor :		**Discharge Nozzle [...]** :	
Tension [V] :	380	**Cooled Bearings** :	**Yes / No**
Rotational direction :	**Clock**	**Cooled Stuffing Box** :	**Yes / No**
Foundation Plate :	**Combined**	**Smothering Gland** :	**Yes / No**
		If yes	
Flexible Coupling :	Yes	**- Seal Liquid** :	**Yes / No**
Pressure Gauge Suction :	No	**- Splash Rings** :	**Yes / No**
Pressure Gauge Discharge :	Yes	**- Packing Type** :	
Min. Overpressure above		**- Mechanical Seal** :	**Yes / No**
p_v/p_m **[bar]** :	0.1	**- N.P.S.H. [m]** :	**9.7**
		$\{=p_m \cdot \rho g\}$	
Construction Materials *(2)*			
Pump House :	MS	**Wear Rings** :	
Pump Rotor :	HT Steel	**Shaft Box** :	
Shaft :	HT Steel		
Special provisions :	none		
Operating Pressure [bara] :	1	**Test Pressure [bara]** :	

Remarks:

(14) Double mechanical seals and seal fluid required for LPG service. Further details to be specified by Rotating Equipment specialist.

MS = Mild Steel; HT Steel = High Tensile Steel

CENTRIFUGAL PUMP - SPECIFICATION SHEET

EQUIPMENT NUMBER	:	**P110**	**Operating**	:	1
NAME	:	**V103 Withdraw Pump**	**Installed Spare**	:	1

Service : **V103 Withdraw Pump**
Type : Centrifugal
Number : 2

Operating Conditions & Physical Data

Pumped liquid			:	Water
Temperature	(T)	**[°C]**	:	37
Density	(ρ)	**[kg/m^3]**	:	1000
Viscosity	(η)	**[N·s/m^2]**	:	6.276E-4
Vapour Pressure	(p_v)	**[bara]**	:	0.063
		at Temperature [°C]	:	37

Power

Capacity	(Φ_v)	**[m^3/s]**	:	9.933E-3
Suction Pressure	(p_s)	**[bara]**	:	1
Discharge Pressure	(p_d)	**[bara]**	:	1.8
Theoretical Power		**[kW]**	:	0.066
Pump Efficiency		**[-]**	:	0.67
Power at Shaft		**[kW]**	:	0.098

Construction Details *(1)*

RPM		:	900	**Nominal diameter**			
Drive		:	Electrical	**Suction Nozzle**	[...]	:	
Type electrical motor		:		**Discharge Nozzle**	[...]	:	
Tension	[V]	:	380	**Cooled Bearings**		:	**Yes / No**
Rotational direction		:	**Clock**	**Cooled Stuffing Box**		:	**Yes / No**
				Smothering Gland		:	**Yes / No**
Foundation Plate		:	**Combined**	**If yes**			
/				**- Seal Liquid**		:	**Yes / No**
Flexible Coupling		:	Yes	**- Splash Rings**		:	**Yes / No**
Pressure Gauge Suction		:	No	**- Packing Type**		:	
Pressure Gauge Discharge		:	Yes	**- Mechanical Seal**		:	**Yes / No**
Min. Overpressure above				**- N.P.S.H.**	**[m]**	:	**3.0**
p_v/p_m	**[bar]**	:	0.1	$\{=p_m \cdot \rho g\}$			

Construction Materials *(2)*

Pump House		:	MS	**Wear Rings**		:	
Pump Rotor		:	HT Steel	**Shaft Box**		:	
Shaft		:	HT Steel				
Special provisions		:	none				
Operating Pressure	**[bara]**	:	1	**Test Pressure**	**[bara]**	:	

Remarks:

(15) Double mechanical seals and seal fluid required for LPG service. Further details to be specified by Rotating Equipment specialist.

MS = Mild Steel; HT Steel = High Tensile Steel

EQUIPMENT NUMBER : **P111**
NAME : **Feeding Wash Water Pump**

Operating : 1
Installed Spare : 1

Service : **Feeding WashWater Pump**
Type : Centrifugal
Number : 2

Operating Conditions & Physical Data

Pumped liquid : Water
Temperature (T) **[°C]** : 25
Density (ρ) **[kg/m^3]** : 1000
Viscosity (η) **[N·s/m^2]** : 7.64E-4
Vapour Pressure (p_v) **[bara]** : 0.032
at Temperature [°C] : 25

Power

Capacity (Φ_v) **[m^3/s]** : 0.028
Suction Pressure (p_s) **[bara]** : 1
Discharge Pressure (p_d) **[bara]** : 4.5

Theoretical Power **[kW]** : 7.1

Pump Efficiency **[-]** : 0.68
Power at Shaft **[kW]** : 10.1

Construction Details *(1)*

RPM : 900
Drive : Electrical
Type electrical motor :
Tension [V] : 380
Rotational direction : **Clock**

Foundation Plate : **Combined**
Flexible Coupling : Yes
Pressure Gauge Suction : No
Pressure Gauge Discharge : Yes
Min. Overpressure above
p_v/p_m [bar] : 0.1

Nominal diameter
Suction Nozzle [...] :
Discharge Nozzle [...] :
Cooled Bearings : **Yes / No**
Cooled Stuffing Box : **Yes / No**
Smothering Gland : **Yes / No**
If yes
- Seal Liquid : **Yes / No**
- Splash Rings : **Yes / No**
- Packing Type :
- Mechanical Seal : **Yes / No**
- N.P.S.H. [m] : **9.9**

Construction Materials *(2)*

Pump House : MS
Pump Rotor : HT Steel
Shaft : HT Steel
Special provisions : none

Operating Pressure [bara] : 1

Wear Rings :
Shaft Box :

Test Pressure [bara] :

Remarks:

Double mechanical seals and seal fluid required for LPG service. Further details to be specified by Rotating Equipment specialist.MS = Mild Steel; HT Steel = High Tensile Steel

VESSEL - SPECIFICATION SHEET

EQUIPMENT NUMBER :	**R101**	**In Series** :	1
NAME :	**Fermentor**	**In Parallel** :	none

General Data	
Service	: Reaction
Type	: - Stirred tank
Position	: - Vertical
Internals	: - Coil
Heating/Cooling medium	: - Cooling water
Vessel Diameter (ID) [m]	: 5.3
Vessel Height [m]	: 9.3
Vessel Tot. Volume [m^3]	: 204.2
Vessel Material	: SS (1)

Process Conditions	
Operational Details	**Fermentation**
Press./Temp.'s [atm/°C]	1/37
Stirring [rpm] :	79
Heating/Cooling :	**Cooling (cooling water T = 10°C)**
Aeration medium :	Mixing gas (2).
Content	Biomass, Glucose, Tyrosine, L-phenylalanine, Water, Acetate, Salts, Ammonia
Vessel Features	
Aeration sparger Stirrer 30-300 RPM Temperature sensor Level indicator pH-sensor Dissolved oxygen level indicator	Cooling coil: D = 0.168 m (3) L = 106 m (4)

Remarks:

1. SS = Stainless Steel
2. 32% pure O_2 and 68% air.
3. Coil pipe diameter
4. Length of coil

Designers :	J. Nie J.F.S.Vasquez M.H.Ucisik H.Huiling	Project ID-Number : Date :	**CPD3324** October 14th 2005

DRYER – SPECIFICATION SHEET

EQUIPMENT NUMBER :	**D 101**	**In Series**	**:** 1
NAME :	**Dryer**	**In Parallel**	**:** none

General Data

Service : - Separation

Type : - Pneumatic Dryers

Position : - Vertical

Internals : - none

Heating/Cooling medium : - hot air

- Type : n.a.

- Quantity [m^3/s] : 50

- Temp. [^oC] : 125

Vessel Diameter (ID) [m] : n.a

Vessel Height [m] : n.a

Vessel Tot. Volume [m^3] : 228

Vessel Material : n.a.

Gas velocity [m/s] : 5.2

Process Conditions

Stream Data		**Feed**		**Product**	
Temperature	**[^oC]**	: 25		: n.a.	
Pressure	**[bara]**	: n.a.		: n.a.	
Density	**[kg/m^3]**	: 1160 (Approx.)		: 1290	
Mass Flow	**[kg/hr]**	: 225.00		: 136.07	
Composition		mol%	wt%	mol%	wt%
L-Phe crystal			60		98.8
Impurities (1)			3.9		1.2
Water			37.5		0 (2)

Remarks:

(1) The impurities include glucose, acetate, and etc.

(2) Assume all the water is evaporated in the dryer.

Designers :	J. Nie J.F.S.Vasquez M.H.Ucisik H.Huiling	Project ID-Number :	**CPD3324**
		Date :	October 14th 2005

FILTER - SPECIFICATION SHEET

EQUIPMENT NUMBER :	**S 102**	**In Series :**	1
NAME :	**Filter**	**In Parallel :**	2

General Data

Service		: Separation
Type		: - Plate and frame filters
Position		: - Horizontal
Internals		: - none
Heating/Cooling medium		: - none
Plates size	[m×m]	: 1.5×1.7
Total filtration area	[m^2]	: 30.6
Pore size	[μm]	: 10
Number of plates		: 6
Frame thickness	[m]	: 0.2
Vessel Material		: Cast iron

Process Conditions

Stream Data	**Feed**		**Retentate (1)**		**Permeate**	
Temperature [°C]	: 45		: 45		: 45	
Pressure[bara]	: 1 to 4.36		: 1		: 1	
Density[kg/m^3]	: 1160		: 1290		: 1000	
Mass Flow[kg/s]	: 8.40		: 0.06		: 8.34	
Composition	mol%	wt%	mol%	wt%	mol%	wt%
L-Phe		**4.4**		**61.2**		**4.0**
Glucose		**1.6**		**0.6**		**1.6**
Acetate		**0.7**		**0.3**		**0.7**
Water		**93.3**		**37.5**		**93.7**

Remarks:

(1) This retentate is the amount before washing.

Designers :	J. Nie J.F.S.Vasquez M.H.Ucisik H.Huiling	Project ID-Number : Date :	**CPD3324** October 14th 2005

VESSEL – SPECIFICATION SHEET

EQUIPMENT NUMBER :	**T101**	**In Series**	: 1
NAME :	**Glucose Storage Tank**	**In Parallel**	: none

General Data

Service		:	Storage
Type		:	- Cylindrical
Position		:	- Vertical
Internals		:	-none
Heating/Cooling medium		:	-none
-Type		:	n.a.
-Quantity	**[kg/s]**	:	n.a.
-Press./Temp.'s	**[bara/°C]**	:	n.a.
Vessel Diameter	**[m]**	:	6.1
Vessel Height		:	12
Vessel total volume	**[m3]**	:	358.8
Vessel Material		:	SS

Process Conditions

Temperature	**[°C**	:	25
Pressure	**[bara]**	:	1
Contents			Glucose Water
Special features			Pressure sensor +emergency valve Level indicator

Remarks:
(1) SS=Stainless Steel

Designers :	J. Nie J.F.S.Vasquez M.H.Ucisik H.Huiling	Project ID-Number : **CPD3324** Date : October 11th 2005

VESSEL - SPECIFICATION SHEET

EQUIPMENT NUMBER :	**T102**	**In Series:**	1
NAME :	**Tyrosine Storage Tank**	**In Parallel:**	none

General Data

Service		: - Storage
Type		: - Cylindrical
Position		: - Vertical
Internals		: -none
Heating/Cooling medium		: -none
-Type		: n.a.
-Quantity	**[kg/s]**	: n.a.
-Press./Temp.'s	**[bara/ °C]**	: n.a.
Vessel Diameter	**[m]**	: 4.2
Vessel Height		: 8.4
Vessel total volume	**[m3]**	: 116.9
Vessel Material		: SS

Process Conditions

Temperature	**[°C] :**	25
Pressure	**[bara] :**	1
Contents		Tyrosine Water
Special features		Pressure sensor +emergency valve Level indicator

Remarks:

(1) SS=Stainless Steel

Designers : J. Nie J.F.S.Vasquez M.H.Ucisik H.Huiling	Project ID-Number : **CPD3324** Date : October 11th 2005

VESSEL – SPECIFICATION SHEET

EQUIPMENT NUMBER :T103 **In Series** :1
NAME **:Ammonia Water Storage Tank** **In Parallel :** none

General Data

Service : - Storage

Type : - Cylindrical

Position : - Vertical

Internals : --none
Heating/Cooling medium : -none
-Type : n.a.
-Quantity **[kg/s]** : n.a.
-Press./Temp.'s **[bara/ °C]** : n.a.
Vessel Diameter **[m]** : 2.5
Vessel Height : 5
Vessel total volume **[m3]** : 24.3
Vessel Material : SS

Process Conditions

Temperature **[°C]** :	25
Pressure **[bara]** :	1
Contents	Ammonia Water
Special features	Pressure sensor +emergency valve Level indicator

Remarks:
(1) SS=Stainless Steel

Designers : J. Nie J.F.S.Vasquez M.H.Ucisik H.Huiling	Project ID-Number : **CPD3324** Date : October 11th 2005

APPENDIX.7 MASS AND ENERGY BALANCES

Appendix7.1 Heat and Mass Balances for Total

HEAT & MASS BALANCE FOR STREAMS TOTAL

IN						OUT				
		EQUIPMENT			EQUIPM.	EQUIPMENT			Plant	
Mass kg/h	Heat kW	Mass kg/h	Heat kW	Stream Nr.	IDENTIF.	Stream Nr.	Mass kg/h	Heat kW	Mass kg/h	Heat kW
1379.76	**-3085.67**	1379.76	-3085.67	**<102>**	**R101**	**<111>**	2346.54	-1772.29	**2346.54**	**-1772.29**
347.90	**-1272.74**	347.90	-1272.74	**<104>**		**<114>**	62687.05	-224106.21	**62687.05**	**-224106.21**
72.35	**-215.25**	72.35	-215.25	**<106>**						
2122.42	**-2.90**	2122.42	-2.90	**<110>**						
10447.68	**-33142.35**	10447.68	-33142.35	**<116>**						
15859.14	**-56079.68**	15859.14	-56079.68	**<133>**						
34800.00	**-129340.00**	34800.00	-129340.00	**<139>**						
4.34	**0.00**	65029.25	-223138.59		Total		65033.59	-225878.50	**0.00**	**2739.91**
62687.05	**-224106.21**	62687.05	-224106.21	**<115>**	**S101**	**<116>**	10447.68	-33142.35	**10447.68**	**-33142.35**
						<117>	52233.38	-190942.01	**52233.38**	**-190942.01**
0.00	**21.85**	62687.05	-224106.21		Total		62681.05	-224084.36	**6.00**	**0.00**
52233.38	**-190942.01**	52233.38	-190942.01	**<119>**	**V101**	**<121>**	30239.82	-106511.38	**30239.82**	**-106511.38**
13766.95	**-48681.46**	13766.95	-48681.46	**<132>**		**<135>**	35760.00	-108770.00	**35760.00**	**-108770.00**
0.00	**24342.09**	66000.32	-239623.47		Total		65999.82	-215281.38	**0.50**	**0.00**
30239.82	**-106511.38**	30239.82	-106511.38	**<122>**	**V102**	**<123>**	30231.13	-106480.77	**30231.13**	**-106480.77**
0.00	**30.61**	30239.82	-106511.38		Total		30231.13	-106480.77	**8.69**	**0.00**
30231.13	**-106480.77**	30231.13	-106480.77	**<124>**	**S102**	**<125>**	225.00	-388.63	**225.00**	**-388.63**
						<130>	30010.04	-106118.84	**30010.04**	**-106118.84**
3.91	**0.00**	30231.13	-106480.77		Total		30235.04	-106507.47	**0.00**	**26.70**
254195.83	**-875466**				Total				**254195.83**	**-875466**
OUT - IN :									0.00	**0**

Project ID Number : **CPD3324**
Completion Date : 16-10-2005

Appendix 7.2 Component Balance for the Production Phase

Overall Component Mass Balance & Stream Heat balance for production phase							
STREAM Nr. :		<101><103><105>,<109>	IN	<113><140><134><125>	OUT		OUT-IN
	Name :	Total Plant		Total Plant		Total Plant	
COMP	MW	kg/hr	kmol/hr	kg/hr	kmol/hr	kg/hr	kmol/hr
Glucose	180.16	965.83	5.3610	7.61	0.0422	-958.22	-5.3187
Tyrosine	181.19	8.70	0.0480	0.00	0.0000	-8.70	-0.0480
L-Phe	165.19	0.00	0.0000	153.89	0.9316	153.89	0.9316
Acetate	59	0.00	0.0000	3.37	0.0572	3.37	0.0572
NH4	18	18.1	1.0049	0.08	0.0043	-18.01	-1.0006
H2O	18	807.4	44.8553	1410.94	78.3856	603.55	33.5303
O2	32	488.3	15.2579	0.10	0.0032	-488.15	-15.2547
CO2	44	1.2	0.0265	713.43	16.2144	712.27	16.1878
N2	28	1633	58.3214	1633.00	58.3214	0.00	0.0000
Total		3922.43	124.8750	3922.42	153.9599	-0.01	29.0850
Enthalpy	kW	-4577		-7112		-2536	

APPENDIX 8 PROCESS AND EQUIPMENT DESIGN

Appendix 8.1 Fermentor

A standard stirred tank is used in the process and designed as follows.

The normal value of power input by stirring in the industry is 2000W/m^3.

$P\_ov := 2000$

Density of the broth is assumed to be equal with the water. [kg/m^3]

$\rho_w := 1000$

The liquid volume (m^3) in the fermentor is obtained from the mass balance calculation. $Vl := 116$

Assume the diameter (D) of the fermentor is equal to the liquid height (H) in the fermentor, therefore,

$$\begin{bmatrix} D \\ H \end{bmatrix} = \begin{bmatrix} 5.286 \\ 5.286 \end{bmatrix}$$

For our process, the dissolved oxygen concentration in the fermentor should maintain in 40%. There are three ways to increase the oxygen transfer amount: 1) increase the gas flow rate, 2) increase the pressure in the fermentor, and 3) increase the oxygen concentration in the aeration gas. Since the high gas flow rate may increase the gas hold up in the fermentor or may cause flooding, method one is not used. For method two, although dissolved oxygen concentration can be increase very much by the pressure raising, the dissolved carbon dioxide also increase which is not good for the biomass growth. In conclusion, mixing gas with pure oxygen and pure oxygen can be used in this process.

40% Oxygen in the gas is chosen, assume this concentration does not affect biomass growth.

$xoin := 0.40$	Oxygen fraction in gas
$vs := 0.04$	Superficial gas velocity
$x\_o := 0.2$	Initial oxygen fraction in the off gas
$P\_atm := 1.01325 \cdot 10^5$	Pressure in the top of fermentor. [Pa]
$P_{bottom} := P\_atm + \rho_w \cdot g \cdot H$	Pressure in the bottom of fermentor. [Pa]
$P_f := \frac{(P\_atm + P_{bottom})}{2}$	Assume the pressure in the fermentor is the average pressure of the top and bottom of the liquid
$m_o := 32$	Herry coefficient for oxygen [-]
$T\_f := 310$	Temperature in the fermentor [K]
$OUR := 130.38$	Oxygen uptake rate, [molO2/m3/h], from mass balance calculation
$Col := 0.10$	Dissolved oxygen concentration in the fermentor. [Mol/m3]

$$3600\frac{P\_atm}{R\cdot T\_atm}\cdot\pi\cdot\left(\frac{D}{2}\right)^{2}\cdot vs\cdot(xoin - x) = OUR\cdot Vl$$

$$2.6\cdot 10^{-2}\cdot(P\_ov)^{0.4}\cdot vs^{0.5}\cdot 3600\cdot\left[\left(x\cdot\frac{P_f}{m_o\cdot R\cdot T\_f}\right) - Co\right] = OUR$$

$$\begin{pmatrix} vs \\ x\_o \end{pmatrix} = \begin{pmatrix} 0.04 \\ 0.283 \end{pmatrix}$$

$kla := 0.026 P\_ov^{0.4} vs^{0.5}$

Oxygen transfer coefficient. [1/s].
Equation (11.40) (p.251, Klaas van't Riet, Johannes Tramper. Basic Bioreactor Design)

$\phi gas := vs\cdot\pi\cdot\left(\frac{D}{2}\right)^{2}$

Gas flow rate

$\phi gas = 0.877$

[m^3/s]

Calculation pure oxygen fraction of the aeration gas

$$\frac{\phi_{o2} + (\phi gas - \phi_{o2})\cdot 0.21}{\phi_{o2} + (\phi gas - \phi_{o2})} = xoir$$

$\phi_{o2} = 0.211$ Pure oxygen gas flow rate [m^3/s]

$\phi air = 0.666$ Air flow rate [m^3/s]

$Co\_star := x\_o\cdot\frac{P_f}{m_o\cdot R\cdot T\_f}$

Oxygen saturation concentration in the fermentor [mol/m^3]

$\varepsilon g := 0.13 P\_ov^{\frac{1}{3}}\cdot vs^{\frac{2}{3}}$ Gas hold up in the stirred vessel

$\varepsilon g = 0.191$

$V_{total} := \frac{V}{0.7}$ Assume working volume is 70% of the fermentor

$V_{total} = 204.938$ Total volume of the fermentor [m^3]

$H_{total} := \frac{V_{total}}{\pi\cdot\left(\frac{D}{2}\right)^{2}}$ Total height of the fermentor

$H_{total} = 9.339$ [m]

$Pgov := g\cdot vs\cdot\rho_w$ Power input of gas flow [W/m^3]

$Pg := Pgov\cdot V$ Power input of gas flow [W]

$Ps := P\_ov\cdot V$ Power input of stir [W]

$Pg = 4.541\times 10^{4} < Ps = 2.32\times 10^{5}$ The gas flow rate does not cause flooding

$Np := 6$

$Dstir := \frac{D}{3}$ Diameter of the stir

$N := \left(\frac{Ps}{Np \cdot \rho_w \cdot Dstir^5}\right)^{\frac{1}{3}}$ Fermentation technology handouts

$N \cdot 60 = 78.933$ [RPM]

$\tau m := \frac{\alpha}{\beta^{\frac{4}{3}} \cdot \gamma^{\frac{1}{3}}} \cdot \left(\frac{Ls}{H}\right)^2 \cdot \left(\frac{H}{D}\right)^2 \cdot \frac{D^{\frac{2}{3}}}{(P\_ov + Pgov)^{\frac{1}{3}}}$ Characteristic time for mixing [s]. (Fermentation technology handouts.)

$\tau oc := \frac{Co\_star \cdot 3600}{OUR}$ Characteristic time for oxygen consumption [s]

$\tau olt := \frac{1}{kla}$ Characteristic time for oxygen transfer [s]

$\tau m = 3.7 < \tau olt = 9.202 < \tau oc = 12.046$ So there is no oxygen limitation in this design.

$\tau_{gluc} := \frac{Cglu}{(q_{Glu} \cdot Cx)}$

Characteristic time for glucose consumption [s], where, C_{glu} and Cx are the glucose and biomass concentration in the fermentor.

$\tau_{gluc} = -2.879 \times 10^3\ s$ It is much smaller than the mixing time, so there is no glucose mixing limitation

Utility calculation of the fermentor

$cpg := 130($ Thermal coefficient of the gas. [J/kg/°C]

$\rho gas := 1.3$ Gas density [kg/m^3]

$cv := 2350$ Heat of vaporization [J/kg/°C]

Equilibrium water vapor concentration at broth temperature and top pressure [kg/m^3]

$\rho vo := 0.05$

$\rho vi := 0$ Water vapor concentration of the inflowing air. Assume it is dry air

$hw := 10$ Total (mainly air side) heat transfer coefficient [W/m^2/°C]

$\alpha f := \pi \cdot \left(\frac{D}{2}\right)^2 + \pi \cdot D \cdot H$ Fermentor surface area [m^2]

The formulas used below are in (Klaas van't Riet, Johannes Tramper. Basic Bioreactor Design, Chapter 13)

Heat dissipated by the stirrer [W]

$rhs := Ps$ $\qquad rhs = 2.32\times 10^{5}$

Heat generated by the through flowing gas [W]

$rhg := cpg\cdot(T\_atm - T\_f)\cdot\phi gas\cdot\rho gas$

$rhg = -1.778\times 10^{4}$

Heat generated by the metabolic activity of the microorganisms [W]

$rhm := 460\cdot 10^{3}\cdot\frac{OUR}{3600}\cdot V$ $\qquad rhm = 1.933\times 10^{6}$

Heat loss resulting from vaporization [W]

$rhv := -cv\cdot\phi gas\cdot\left(\rho vo\cdot\frac{P\_atm}{P_f} - \rho vi\right)$ $\qquad rhv = -82.031$

Heat loss through the fermentor walls [W]

$rhw := -hw\cdot\alpha f\cdot(T\_f - T\_atm)$ $\qquad rhw = -1.317\times 10^{4}$

Neglect the effect of feed streams and out stream of the fermentor on the heat calculation, therefore the total heat of the fermentor is

$rht := rhs + rhg + rhm + rhv + rhw$

$rht = 2.134\times 10^{6}$ [W]

Assume the broth has the characteristics of water

$\lambda_b := 0.607$ Heat conductivity of the broth, [W/(m*K)]

$\nu_b := 10^{-6}$ Kinematic viscosity of the broth [m^2/s]

$\eta_b := 6.29\cdot 10^{-4}$ Dynamic viscosity of the broth. [Ns/m^2]

$c_{pb} := 4200$ Thermal coefficient of the broth [J/kg/K]

$d_w := 5\cdot 10^{-3}$ Wall thickness cooling coil [m]

$\lambda_w := 8C$ Heat conductivity of the well [W/m/K]

No fouling layer is present and the heat transfer coefficient of the cooling medium can be neglected.

$$\frac{h_l\cdot T_v}{\lambda_b} = 0.6\cdot\left(\frac{N\cdot D_s^{2}}{\nu_b}\right)^{0.67}\cdot\left(\frac{c_{pb}\cdot\eta_b}{\lambda_b}\right)^{0.33}$$

$h_l = 3.009\times 10^{3}$ Heat transfer coefficient medium (fermentation liquid) [$W/m^2/K$]

$$\frac{1}{h_t} = \frac{1}{h_l} + \frac{d_w}{\lambda_w}$$

$h_t = 2.533\times 10^3$	Total heat transfer coefficient. [W/m^2/K]
$T_{ci} := 28$	The cooling water is 10°C. [K]
$T_{co} := 273 + 3$	Assume the temperature of the outflow of the cooling water [K]
$T_b := 273 + 37$	Broth temperature [K]
$H_c := rht$	Heat balance of the fermentor and the cooling
$H_c = h_t \cdot \alpha_c \cdot \dfrac{T_{co} - T_{ci}}{\ln\left(\dfrac{T_b - T_{ci}}{T_b - T_{co}}\right)}$	
$\alpha_c = 56.856$	Surface area of cooling. [m^2]
$d_{c\_chosen} := 0.168$	Chosen cooling coil pipe: Pipe size: in.6; schedule number: 40; Out diameter: 6.625 in
$L_c := \dfrac{\alpha_c}{\pi \cdot d_{c\_chosen}}$	Length of the cooling coil
$L_c = 107.726$	[m]
$Cp_w := 4182 \dfrac{J}{kg \cdot K}$	Heat capacity of water, calculated in the pure component properties.
$rht \cdot W = F_w \cdot Cp_w \cdot (T_{co} - T_{ci}) \cdot K$	
$F_w = 9.183\times 10^4 \dfrac{kg}{hr}$	Cooling water flow rate

Appendix 8.2 Centrifuge

The selection of the biomass recycle equipment was made using Graph 10.10 from Coulson & Richardson's page 407.

After deciding to use a centrifuge, table 18-12 and 18-13 in Perry's Chemical Engineer's Handbook was used to determine the type needed taking in account the steam properties The detailed characteristics of the centrifuge were taken from results of similar applications listed in table 2 and 4 from the Bioseparations handbook.

The power required for the operation of the centrifuge is listed in Table 18-12 from the Perry's Chemical Engineer's Handbook.

Appendix 8.3 Evaporator

The chosen equipment for evaporator is vertical tube forced circulation evaporator. Pumping to cause forced circulation of the liquid inside the tubes can increase the liquid-film heat-transfer coefficient. This can be done in the long-tube vertical type by adding a pipe connection with a pump between the outlet concentrate line and feed line.

The normal practice in the design of forced-convection evaporator is to calculate the heat-transfer coefficient assuming that the heat is transferred by forced convection only. The heat-transfer coefficient can be evaluated using correlations for convective boiling, such as Chen's method. According to Coulson and Richardson conventional shell and tube exchanger designs can be used for the design of forced-circulation evaporators.

Therefore, in this design conventional shell and tube passes are used, with one-shell pass and eight-tube passes where the process liquid is on the shell side. Evaporator has a heat duty of 24270 kW (8.737×10^7 kJ/hr). The amount of water to be evaporated is 3.576×10^4 kg/hr. Saturated steam at atmospheric pressure is used for this duty. The steam totally condensed during the operation and leaves the stream as liquid at 100°C.

Table A8.1 Stream In and Out Temperatures

	In (°C)	Out (°C)
Process Liquid:	38.67	45
Steam:	100	100

The required amount of steam is calculated in the following way:

Table A8.2 Required Steam Mass Flow Calculation

ΔH (kJ/kg)	Heat Duty (kJ/hr)	Steam (kg/hr)
2395	8.737×10^7	38710

Calculations have been done in MathCad program and its notation is used during the explanations.

Process flow stream = Outlet Stream of Centrifuge + Recycle Stream to Evaporator

= 52236 kg/hr + 13770 kg/hr

Process Flow Stream = 66010 kg/hr

- Fraction of vapor:

$$x := \frac{W}{F_{IN}} \qquad x = 0.542$$

- Calculation of the inlet temperature:

$$T_1 := \frac{5.2236\times 10^4 \cdot 37 + 1.377\times 10^4 \cdot 45}{F_{IN}}$$

- Calculation of log-mean temperature difference:

$T_1 = 38.669$ ^oC Inlet Shell Side Fluid Temperature

$T_2 := 45$ ^oC Outlet Shell Side Fluid Temperature

$$\Delta T_{Lm} := \left[\frac{T_2 - T_1}{\ln\left[\frac{T_S - T_1}{T_S - T_2}\right]}\right]$$

$\Delta T_{Lm} = 58.108$ ^oC

$\Delta T := \Delta T_{Lm}$

$q = U \cdot A \cdot \Delta T$

The overall heat-transfer coefficient in an evaporator is composed of the steam-side condensing coefficient. According to Table 8.3-1 in Geankoplis (page 496, in third edition) the value of U may vary between 2300-11000 W/m^2*K for Long-tube vertical forced circulation type of evaporators.

$U := 4000 \; \frac{W}{m^2 \cdot C}$ First initial value for overall heat transfer coefficient

- Area

$A := \frac{Qevap}{U \cdot \Delta T}$ $A = 375.895\ m^2$

Steam will flow through the tubes.

- Tubes:

$Do := 19.05\,10^{-3}$ m Outside Diameter

$Di := 14.83\,10^{-3}$ m Inside Diameter

Allowing for tube-sheet thickness, take $L := 8$ m

Number of passes $Pass := 8$

Area of 1 tube $A_{tube} := \pi \cdot Do \cdot L$ $A_{tube} = 0.479$ m^2

Number of Tubes (Nt) $Nt := \frac{A}{A_{tube}}$ $Nt = 785.111$

For each passes $Nt_{perpass} := \frac{Nt}{Pass}$ $Nt_{perpass} = 98.139$

- Tube Side Velocity

Tube cross sectional area $A_{cross} := \frac{\pi}{4} \cdot Di^2 \cdot Nt$ $A_{cross} = 0.136$ m^2

Area per pass $A_{perpass} := Nt_{perpass} \cdot A_{cross}$ $A_{perpass} = 13.309$ m^2

$\rho\ steam := 0.596$ kg/m^3

Volumetric Flow in Tube $v_{vol} := \frac{S}{3600} \cdot \frac{1}{\rho steam}$ $v_{vol} = 18.041$ $\frac{m^3}{s}$

Tubeside velocity $u_t := \frac{v_{vol}}{A_{perpass}}$ $u_t = 1.356$ $\frac{m}{s}$

The velocity is satisfactory between 1 to 2 m/s.

- Bundle and Shell Diameter

Triangular square patterns are chosen due to its higher transfer rate:

Constants K1 and n1 (Table 12.4, page 649):

Table 12.4 (Sinnott p.649)

$K_1 := 0.0365$

$n_1 := 2.675$

$$D_b := Do \cdot \left(\frac{Nt}{K_1}\right)^{\frac{1}{n_1}}$$

$D_b = 0.794$

Use a Split-Ring Floating Head Type

From Figure 12.10(p.647), Bundle Diameter Clearance: $D_c := 0.06$

Shell Inside Diameter, Ds: $D_s := D_b + D_c$ $D_s = 0.861$

Number of tubes in centre row:

Triangular pitch $P_t := 1.25 \cdot Do$ $N_r := \frac{D_b}{P_t}$ $N_r = 33.325$ Tubes in centre row

- Tube-Side Heat Transfer Coefficient

$\rho water := 958.3 \quad \frac{kg}{m^3}$ $\qquad \rho steam := 0.59 \quad \frac{kg}{m^3}$

$\mu_{water} := 0.2838 \cdot 10^{-3} \ \frac{Pa}{s}$ $\qquad \mu_{steam} := 1.295\,10^{-5} \quad \frac{Pa}{s}$

$Cp_{water} := 4.219$ kJ/kg*K $\qquad Cp_{steam} := 1.88$ kJ/kg*K

$$\rho_{s\_avg} := \frac{\rho water + \rho steam}{2} \qquad \rho_{s\_avg} = 479.488 \ \frac{kg}{m^3}$$

$$\mu_{s\_avg} := \frac{\mu_{water} + \mu_{steam}}{2} \qquad \mu_{s\_avg} = 1.484 \times 10^{-4} \ \frac{Pa}{s}$$

$$Cp_{s\_avg} := \frac{Cp_{water} + Cp_{steam}}{2} \qquad Cp_{s\_avg} = 3.054 \ \frac{kJ}{kg \cdot K}$$

$Mw_{water} := 18$

- Thermal conductivity (k) and Flow Parameters (Re, Pr)

$$k_{s\_avg} := 3.56\,10^{-5} \cdot Cp_{s\_avg} \cdot \left[\frac{\rho_{s\_avg}{}^{4}}{Mw_{water}} \right]^{\frac{1}{3}} \qquad k_{s\_avg} = 0.156$$

$$Re := \frac{Di \cdot u_t \cdot \rho_{s\_avg}}{\mu_{s\_avg}} \qquad Re = 6.497 \times 10^{4}$$

$$Pr := \frac{Cp_{s\_avg} \cdot \mu_{s\_avg}}{k_{s\_avg}} \qquad Pr = 2.91 \times 10^{-3}$$

$$\frac{L}{Di} = 539.447$$

- Calculation of h_i:

From Figure 12.23 page 665 in Sinnott $j_h := 3.1 \cdot 10^{-3}$

$h_i := 500($

Given $$\frac{h_i \cdot Di}{k_{s\_avg}} = j_h \cdot Re \cdot Pr^{0.33} \cdot \left(\frac{\mu_{s\_avg}}{\mu_{water}}\right)^{0.14}$$

$(h_i) := \text{Find}\,(h_i)$

$h_i = 281.045 \quad \frac{W}{m^2 \cdot C}$

Boyko-Kruzhilin Correlation

$$h_{c\_BK} := h_i \cdot \left[\frac{1 + \sqrt{\left(\frac{\rho water}{\rho steam}\right)}}{2}\right] \qquad h_{c\_BK} = 5.775 \times 10^3 \quad \frac{W}{m^2 \cdot C}$$

$$h_{i\_prime} := h_{c\_BK} \qquad h_{i\_prime} = 5.775 \times 10^3 \quad \frac{W}{m^2 \cdot C}$$

Shell-side Heat Transfer Coefficient

- Baffle spacing

Baffle spacing should give good transfer without too high pressure drop. The used range is from 0.2 to 1.0 shell diameter. A close baffle spacing will give higher heat transfer coefficients but at the expense of higher pressure drop.

$l_B := D_s \cdot 0.8 \quad l_B = 0.668 \quad m$

- Equivalent diameter

$$d_e := \frac{1.27}{Do} \cdot \left(P_t^2 - 0.785 Do^2\right) \qquad d_e = 0.019$$

$\rho_{shellliquid} := 1000\ \frac{kg}{m^3}$

$\mu_{water\_mean} := \left(\frac{0.6097 + 0.5883}{2}\right) \cdot 10^{-3}$

$Cp := 4.18$

$k_s := 3.56\,10^{-5} \cdot Cp \cdot \left(\frac{\rho_{shellliquid}}{Mw_{water}}^{4}\right)^{\frac{1}{3}}$ $\qquad k_s = 0.568\ \frac{W}{m \cdot C}$

- Volumetric flow rate on shell side

$v_{shell\_vol} := \frac{F_{IN}}{3600} \cdot \frac{1}{\rho_{shellliquid}}$ $\qquad v_{shell\_vol} = 0.018\ \frac{m^3}{s}$

$u_s := \frac{v_{shell\_vol}}{A_s}$ $\qquad u_s = 0.155\ \frac{m}{s}$

$Re_{shell} := \frac{d_e \cdot u_s \cdot \rho_{shellliquid}}{\mu_{water}}$ $\qquad Re_{shell} = 1.026 \times 10^{4}$

$Pr_{shell} := \frac{Cp \cdot \mu_{water\_mean}}{k_s}$ $\qquad Pr_{shell} = 4.41 \times 10^{-3}$

Figure 12.29 p.672, take 25% baffle cuts, which is a reasonable guess

$j_{h\_shell} := 5.7 \cdot 10^{-3}$

$h_s := 5000$

Given

$$\frac{h_s \cdot d_e}{k_s} = j_{h\_shell} \cdot Re_{shell} \cdot Pr_{shell}^{\frac{1}{3}} \cdot \left(\frac{\mu_{water\_mean}}{\mu_{water}}\right)^{0.14}$$

$(h_s) := \text{Find}\,(h_s)$

$h_s = 321.288\ \frac{W}{m^2 \cdot C}$

- Lockart-Martinelli two-phase flow parameter

$X_{tt} := 1$

Given

$$\frac{1}{X_{tt}} = \left(\frac{x}{1-x}\right)^{0.9} \cdot \left(\frac{\rho water}{\rho steam}\right)^{0.5} \cdot \left(\frac{\mu_{steam}}{\mu_{water}}\right)^{0.1}$$

$(X_{tt}) := \text{Find}\,(X_{tt})$ $\qquad X_{tt} = 0.029$

$$\frac{1}{X_{tt}} = 34.24$$

- The two-phase correction factor, fc

From Figure 12.56, p.734 $\quad fc := 33$

- Corrected Forced Convection Coefficient, hsc:

$h_{sc} := h_s \cdot fc$ $\qquad h_{sc} = 1.06\times 10^4 \quad \frac{W}{m^2 \cdot C}$

- The reduced pressure correlation of Mostinski (1963) to estimate nucleate boiling coefficient:

Nucleate boiling coefficient, hnb

$P := 0.09$ bar

$P_c := 220.$ bar

$q := \frac{Qevap}{1000\,A_s}$ (Heat Flux) $\qquad q = 7.374\times 10^5 \quad \frac{W}{m^2 \cdot hr}$

$$h_{nb} := 0.104\left(P_c\right)^{0.69} \cdot (q)^{0.7} \cdot \left[1.8 \cdot \left(\frac{P}{P_c}\right)^{0.17} + 4 \cdot \left(\frac{P}{P_c}\right)^{1.2} + 10 \cdot \left(\frac{P}{P_c}\right)^{10}\right]$$

$h_{nb} = 2.66\times 10^4 \quad \frac{W}{m^2 \cdot C}$

$Re_L := \frac{(1-x) \cdot d_e \cdot u_s \cdot \rho_{shellliquid}}{\mu_{water}}$ Liquid Reynolds number

- Corrected Liquid Reynolds Number

$Re_L \cdot fc = 1.551 \times 10^5$

Figure 12.57, page:735 $f_s := 0.27$ Suppression factor

- Corrected Heat Transfer Coefficient

$h_{nb\_prime} := f_s \cdot h_{nb}$ $h_{nb\_prime} = 7.181 \times 10^3$ $\frac{W}{m^2 \cdot C}$

$h_{sc\_prime} := h_{sc} + h_{nb\_prime}$

$h_{sc\_prime} = 1.778 \times 10^4$ $\frac{W}{m^2 \cdot C}$

- Pressure Drop in Shell-Side

Read friction factor from Figure 12.30, p.673: $j_f := 4.8 \cdot 10^{-2}$

$$\Delta P_s := 8 \cdot j_f \cdot \left(\frac{D_s}{d_e}\right) \cdot \left(\frac{L}{l_B}\right) \cdot \left(\frac{\rho_{shellliquid} \cdot u_s^2}{2}\right) \cdot \left(\frac{\mu_{water\_mean}}{\mu_{water}}\right)^{-0.14}$$

$\Delta P_s = 2.201 \times 10^3$ $\frac{N}{m^2}$

Convert to bar: $\Delta P_s := \Delta P_s \cdot 10^{-5}$ $\Delta P_s = 0.022$ bar Very low pressure drop

- Overall Heat Transfer Coefficient

"The normal practice in the design of forced-circulation evaporators is to calculate the heat-transfer coefficient assuming that the heat is transferred by forced convection only." Coulson and Richardson; (Sinnott, R.K.) Chapter 12.11.4 Design of forced-circulation reboilers, Volume 6, Third Edition, page 737

$U_o := 1000$ W/m*C

Given

$$\frac{1}{U_o} = \frac{1}{h_{sc\_prime}} + \frac{Do}{Di} \cdot \frac{1}{h_{i\_prime}}$$

$$(U_o) := \text{Find}(U_o)$$

$$\boxed{U_o = 3.589 \times 10^3} \quad \text{W/m*C}$$

The calculated overall heat coefficient is close to the assumed one, 4000 W/m*C. That means no further calculation is required.

Appendix 8.4 Crystallizer

The design of the evaporative crystallization device is split in two equipments for calculation purposes, f the evaporation section of the crystallizer is designed and then the crystallizer, the crystallizer will consist on a tank with a cooling jacket that will remove the energy of solution needed to form the crystals. The energy that needs to be removed is negligible in comparison with the energy needed to evaporate the water.

$S := \frac{C_{Phe\_2}}{C_{Phe\_3}}$ Supersaturation

$S := 1.10$ Anhydrate crystals constants

$g := 0.72$ Anhydrate crystals constants

$Kg := 1.141 \cdot 10^{-6} \cdot \frac{m}{s}$ Anhydrate crystals constants

$G := Kg \cdot (S - 1)^{g}$ Crystal growth

$G = 2.247 \times 10^{-7} \frac{m}{s}$

$Lm := 1 \cdot 10^{-4} \cdot m$ Crystal mean size

$\tau := \frac{Lm}{3G}$ Residence time

$\tau = 148.331 s$

$V := \phi out \cdot \tau$ Volume of the crystallizer

$V := 1.245 m^{3}$

$Dcr := \left(\frac{8 \cdot Vc}{3 \cdot \pi}\right)^{\frac{1}{3}}$ Crystallizer diameter

$Dcr := 1.018 m$

$H := \frac{3 \cdot Dcr}{2}$ Crystallizer height

$H := 1.527 m$

Appendix 8.5 Plate Frame Filter

$\phi mc := 135\frac{kg}{hr}$	Crystal in the feed stream to the filter
$\phi w\_in := 2.82\times 10^{4}\frac{kg}{hr}$	Water mass flow rate of the liquid to the filter
$\rho w := 1000\frac{kg}{m^{3}}$	Assume the solution density is the same with water density
$\phi_{vin} := \frac{\phi w\_in}{\rho w}$	Volume flow rate of the liquid to the filter
$\phi mc\_w := \phi mc\cdot\frac{4}{6}$	Assume after filtration, water is 40% of the product. Other component in this stream is neglected.
$\phi_{v\_per} := \phi_{vin} - \frac{\phi mc\_w}{\rho w}$	Permeate flow rate of the filter

Assume use 1.5 by 1.8 m plates

$A_{m} := 1.5m\cdot 1.8m\cdot 11$	Membrane area
$J_{per} := \frac{\phi_{v\_per}}{A_{m}}$	Permeate flux [$m^3/m^2/h$]
$t_{f} := 48hr$	Filtration time
$w(t) := \phi mc\cdot\frac{t}{A_{m}}$	Assume 100% filtration
$\alpha := 10^{10}\frac{m}{kg}$	Specific cake resistance. From bioseperation handouts.
$R_{c}(t) := \alpha\cdot w(t)$	Resistance of deposited layers [1/m]
$\mu_{per} := 0.555\,0.01poise$	Viscosity of permeate, use viscosity of 45°C water

The membrane resistance can be neglected compared with the cake resistance. According to the equation (20.2) (p.873, Coulson & Richardson's, Chemical engineering, volume 2)

$\Delta P(t) := \mu_{per}\cdot\left(R_{c}(t)\right)\cdot J_{per}$	Transmembrane pressure
$m_{w} := \phi mc\_w\cdot t_{f}$	Total water amount contained in the product

Estimate washing time and washing water volume

$R := \frac{1.2\%}{4\%}$	The weight fraction of solute remaining in the cake after washing
$E := 7($	Assume percentage wash efficiency
$R = \left(1 - \frac{E}{100}\right)^n$	Equation (4.2.7) (p.110, Roger G.Harrison, Bioseparations science and engineering)
$n = 1$	The volume of wash liquid per volume of liquid in the unwashed cake
$V_r := \frac{m_w}{\rho w}$	Residual liquid in the cake
$V_f := \phi_{v\_per} \cdot t_f$	Volume of filtrate after time tf
$f := \frac{V_r}{V_f}$	The ratio of V_r and V_f
$t_w := 2 \cdot n \cdot f \cdot t_f$	Equation (4.2.14) (p.111, , Roger G.Harrison, Bioseparations science and engineering)
$V_w := \frac{t_w \cdot V_f}{2 \cdot t_f}$	Washing water volume
$M_w := V_w \cdot \rho w$	Washing water mass for one filter in one filtration batch

$M_w = 4.32 \times 10^3 \, kg$

Appendix 8.6 Compressor

The compressor is selected according to the air flow and the power required for the compression.

Formulas and method from transpor procesess and unit operations (Christie J. Geankoplis)

$Mw := 28.84 \cdot \frac{gm}{mol}$ 1 mol of air wights 28.84

$Cp\_air := Mw \cdot 1.0048 \cdot \frac{J}{gm \cdot K}$

$Cp\_air = 28.978 \frac{J}{mol \cdot K}$

$\rho air := 1.15 \cdot \frac{kg}{m^3}$

$R := 8.3144 \cdot \frac{J}{mol \cdot K}$

$\phi vair := 0.877 \cdot \frac{m^3}{s}$ Gas flow in to the compressor

$\phi vair = 3157.2 \frac{m^3}{hr}$

$\phi air := \phi vair \cdot \rho air$

$\phi air = 3.631 \times 10^3 \frac{kg}{hr}$

$W_{air} := \phi air \cdot 0.03685$

$k := \frac{Cp\_air}{Cp\_air - R}$

$k = 1.402$

$T_s := 298.15 \cdot K$

$Ps := 1 \cdot atm$ Inlet Pressure to the compressor

$Pd := 2.5 \cdot atm$ Discharge Pressure of the compressor

$$Had := \frac{k \cdot \frac{R}{Mw} \cdot T_s}{k-1} \cdot \left[\left(\frac{Pd}{Ps} \right)^{\frac{k-1}{k}} - 1 \right]$$ Perry p.10-31 eq. 10.64a.

$Had = 3.014 \times 10^{4} \frac{ft \cdot lbf}{lb}$ Adiabatic Head

$W\_ad := Had \cdot \phi air$

$W\_ad = 121.836\,hp$ Assuming an efficiency of 80%

$\eta := 80\%$

$Wad := \frac{W\_ad}{\eta}$ $\phi air = 1.009 \frac{kg}{s}$ $Ps = 1\,atm$

$Wad = 152.295\,hp$

$$T_d := T_s \cdot \left(\frac{Pd}{Ps} \right)^{\frac{k-1}{k}}$$

$T_d = 387.802\,K$

METHOD 2

$\gamma air := 1.4$ According to geankoplis

$$Ws := -\left(\frac{\gamma air}{\gamma air - 1} \right) \cdot \frac{T_s \cdot R}{Mw} \cdot \left[\left(\frac{Pd}{Ps} \right)^{\frac{\gamma air - 1}{\gamma air}} - 1 \right]$$

$Ws = -9.003 \times 10^{4} \frac{N \cdot m}{kg}$

$\eta := 80\%$

The break power if the efficiency is 80%:

$bhp := \frac{-Ws \cdot \phi air}{\eta}$ $\phi vair = 3.157 \times 10^{3} \frac{m^3}{hr}$

$bhp = 152.207\,hp$ Adiabatic Compressor

If we use the isothermal compresor:

$$Wiso := -(2.3026)\cdot\frac{T_s\cdot R}{Mw}\cdot\log\left(\frac{Pd}{Ps}\right)$$

$$Wiso = -7.876\times10^4\,\frac{J}{kg}$$

The break power if the efficiency is 80%:

$$bhp_{iso} := \frac{-Wiso\cdot\phi air}{\eta}$$

$bhp_{iso} = 99.292\,kW$ FOR AN ISOTHERMAL COMPRESSOR.

Becouse of the less power rquirements of an isothermal compresor is going to be selected, th cooling in the stages of the compresiion is not taken in consideration

$i_{1998} := 389.5$

$i_{2005} := 465$

$Constant := 960$

Coulson & Richardson's 6th edition

$index := 0.8$

$Size := 99.292$

$$Cost := Constant\cdot Size^{index}$$

$Cost = 3.8\times10^4$ US$

$$Cost\_2005 := Cost\cdot\frac{i_{2005}}{i_{1998}}$$

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$Cost\_2005 = 4.537\times10^4$ US $

$$Cost_{manteinance\_2005} := 0.1\cdot Cost\_2005$$

$$Cost_{manteinance\_2005} = 4.537\times10^3$$

Appendix 8.7 Dryer

According to Coulson and Richardson's Chemical Engineering Vol. 4 , p.731, Table 16.7. Pneumatic Dryers is chosen.

$\phi_{cr} := 135 \frac{kg}{hr}$	L-Phe crystal in the feed stream
$\phi_{w} := 84.33 \frac{kg}{hr}$	Water contained in the product needed to evaporated

Assume all the water can be removed through the dryer

Estimated gas flow rate needed, Because the calculation method for the gas flow rate in the Pneumatic Dryers is not found. The method of estimation of gas flow rate needed to dry a surface is used to roughly estimate the flow rate.

$\phi_{w} := 84.33 \frac{kg}{hr}$	Water contained in the product needed to evaporated
$\tau := 5\,s$	Residence time
$m_s := \phi_{cr} \cdot \tau$	L-Phe crystal suspension mass in the dryer
$r_{cr} := 100 \frac{10^{-6} m}{2}$	Radius of crystal particle
$v_{cr} := \frac{4 \cdot \pi \cdot r_{cr}^{3}}{3}$	Volume of one particle
$\rho := 1290 \frac{kg}{m^3}$	Density of the crystal
$m_{cr} := v_{cr} \cdot \rho$	Mass of one crystal particle
$A_t := \frac{m_s \cdot a_{cr}}{m_{cr}}$	Assumed total surface area of L-Phe crystal in the dryer
$\Delta T := 100\,K$	The drying air temperature is assumed 125°C
$C := 14.5$	Given by SHEPHERD, C.B., HADLOCK, C., AND BREWER, R.C.: Ind.Eng.Chem. 30 (1938) 388. Drying materials in trays. Evaporation of surface moisture.

$\lambda := 2.26 \cdot 10^6 \frac{J}{kg}$ Specific latent heat of vaporization of water

$w = C \cdot G^{0.8} \frac{W}{m^2 K} \cdot A_t \cdot \frac{\Delta T}{\lambda}$ Equation (16.9), (p.695, Coulson and Richardson's Chemical Engineering Vol.4)

$G = 5.989$ Mass flow rate of gas, kg/s/m^2 when dry a flat surface.

The total surface area of crystals is assumed to as the surface area of a flat surface. Therefore,

$M_{gas} := G \cdot \frac{kg}{s \cdot m^2} \cdot A_t$ Mass flow rate of the gas [kg/h]

$\rho_{air} := 1.145 \frac{kg}{m^3}$ Density of the air

$\phi_{v\_air} := \frac{M_{gas}}{\rho_{air}}$ Volume flow rate of the air

$V := \phi_{v\_air} \cdot \tau$ Volume of the dryer

$V = 228.06 \ m^3$

Appendix 8.8 Condensor

The design of the condenser is done according to the procedure described in Coulson and Richardson (Chapter 12.10 Condensers).

The amount of liquid (water steam) to be condensed:

$$W_c := 3.576\times 10^4 \cdot \frac{kg}{hr}$$

The vapor will enter the condensor saturated at 45oC and will leave the system at 37oC as liquid.

Energy required: $\lambda := 2395 \cdot 10^3 \cdot \frac{J}{kg}$ at 45°C and 0.095 bar

$Cp := 4.183\,10^3 \cdot \frac{J}{kg \cdot K}$

$\Delta H := \lambda - Cp \cdot (37 - 45)K$ $\Delta H = 2.428 \times 10^6\ Sv$ $Sv = \frac{J}{kg}$

$Q := W_c \cdot \Delta H$ $Q = 8.684\times 10^{10} \frac{J}{hr}$ $Q = 2.412 \times 10^4\ kW$

Assume the overall heat transfer coefficient $U := 800 \frac{W}{m^2 \cdot K}$

4 tube passes

$Di := 18 \cdot mm$ Inside Diameter of tubes

$Do := 20mm$ Outside Diameter of tubes

$L := 1.83m$ Length of tubes (8 ft)

- Calculation of Logarithmic Mean Temperature Difference

The process liquid is desired to be condensed totally is flowing through the shell sides, and the cooling stream flows through the tube.

$T_1 := (45 + 273.15)K$	Inlet Shell Side Fluid Temperature
$T_2 := (37 + 273)K$	Outlet Shell Side Fluid Temperature
$t_1 := (5 + 273)K$	Inlet tube Side Fluid Temperature
$t_2 := (25 + 273)K$	Outlet Tube Side Fluid Temperature

$$\Delta T_{Lm} := \left[\frac{(T_1 - t_2) - (T_2 - t_1)}{\ln\left(\frac{T_1 - t_2}{T_2 - t_1}\right)}\right]$$

$\Delta T_{Lm} = 25.62K$

- The Correction of Logarithmic Mean Temperature Difference

$R := \frac{T_1 - T_2}{t_2 - t_1}$ $R = 0.407$

$S := \frac{t_2 - t_1}{T_1 - t_1}$ $S = 0.498$

From the Figure 12.19 for one shell pass, 4 tube pass (Coulson and Richardson, 6th Volume, Third Edition).

$F := 0.96$ The correction factor

$\Delta T := \Delta T_{Lm} \cdot F$ $\Delta T = 24.595K$

- Calculation of Heat Transfer Area

$A := \frac{Q}{U \cdot \Delta T}$ $A = 1.226\times 10^3\, m^2$

- Calculation of Amount of Cooling Water Required

Heat Capacity of cold stream: $Cp := 4.183 \cdot 10^3 \cdot \frac{J}{kg \cdot K}$

$F_{tube} := \frac{Q}{(t_2 - t_1) \cdot Cp}$ $F_{tube} = 288.343\, \frac{kg}{s}$ and $F_{tube} = 1.038\times 10^6\, \frac{kg}{hr}$

- Calculation of Number of Tubes (Nt)

$A_{tube} := \pi \cdot Do \cdot L$ $\qquad A_{tube} = 0.115 m^2$

$Nt := \frac{A}{A_{tube}}$ $\qquad Nt = 1.066 \times 10^4$

- Calculation of Bundle and Shell Diameters

Constants K1 and n1:

Table 12.4 (Sinnott p.649)

$K_1 := 0.17$

$n_1 := 2.28$

$$D_b := Do \cdot \left(\frac{Nt}{K_1}\right)^{\frac{1}{n_1}}$$

Equation for square patterns

$D_b = 2.483$ m

Use a Split-Ring Floating Head Type

From Figure 12.10 (p.647), Bundle Diameter Clearance: $D_c := 120$ mm

Shell Inside Diameter, Ds: $D_s := D_b + D_c$ $\qquad D_s = 2.603$ m

Number of tubes in centre row:

Use square pitch $P_t := 1.25 Do$ $\qquad N_r := \frac{D_b}{P_t}$ $\qquad N_r = 99.335$ Tubes in centre row

Calculation of Shell Side Coefficient

Estimate tube wall temperature, Tw; assume condensing coefficient of 700 $W/m^{2} * {}^oC$

$h_c := 700 \frac{W}{m^2 \cdot K}$

Mean temperature: $T_{shell} := \frac{(T_1 + T_2)}{2}$ $\qquad T_{shell} = 314.075 K$

$T_{tube} := \frac{(t_1 + t_2)}{2}$ $\qquad T_{tube} = 288 K$

$T_w := 310K$

Given

$$\left(T_{shell} - T_w\right)\cdot h_c = \left(T_{shell} - T_{tube}\right)\cdot U$$

$T_w := \text{Find}\left(T_w\right)$

$T_w = 284.275\ \text{K}$

Physical properties at: $\frac{T_w}{K} - 273 = 11.275$ ^oC

$\mu_L := 1.26$ $\frac{g}{m\cdot s}$ or $\frac{mNs}{m^2}$

$\rho_L := 999.5$ $\frac{kg}{m^3}$

$M := 18$ $\frac{g}{mol}$

$Cp := 4.19$ $\frac{J}{kg\cdot K}$

$$k_L := 3.56\,10^{-5}\cdot Cp\cdot\left(\frac{\rho_L{}^4}{M}\right)^{\frac{1}{3}}$$

$k_L = 0.569$ $\frac{W}{m\cdot K}$

$\Gamma_h := \frac{W_c}{L\cdot Nt}$ $\quad \Gamma_h = 5.091\times 10^{-4}\ \frac{kg}{ms}$ $\quad \Gamma_h := \Gamma_h\cdot\frac{m\cdot s}{kg}$ $\quad \Gamma_h = 5.091\times 10^{-4}$

$\rho_v := 0.1$ $\frac{kg}{m^3}$

$N_r := \frac{2}{3}\cdot N_t$ $\quad N_r = 66.223$

$$h_c := 0.95 k_L\cdot\left[\frac{\rho_L\cdot\left(\rho_L - \rho_v\right)9.81}{\mu_L\cdot\Gamma_h}\right]^{\frac{1}{3}}\cdot N_r{}^{-\frac{1}{6}}$$

$h_c = 666.532$

$h_c := h_c \cdot \frac{W}{m^2 \cdot K}$

Given

$$\left(T_{shell} - T_w\right) \cdot h_c = \left(T_{shell} - T_{tube}\right) \cdot U$$

$T_w := \text{Find}\ \left(T_w\right)$

$T_w = 282.779 \quad K$

Physical properties at $\frac{T_w}{K} - 273 = 9.779 \quad {}^oC$

$\mu_L := 1.35 \qquad \frac{g}{m \cdot s}$

$\rho_L := 999.7 \qquad \frac{kg}{m^3}$

$Cp := 4.195$

$$k_L := 3.56\,10^{-5} \cdot Cp \cdot \left[\frac{\rho_L{}^4}{M}\right]^{\frac{1}{3}}$$

$k_L = 0.57 \qquad \frac{W}{m \cdot K}$

$\Gamma_h := \frac{W_c}{L \cdot Nt} \qquad \Gamma_h = 5.091 \times 10^{-4} \frac{kg}{ms} \qquad \Gamma_h := \Gamma_h \cdot \frac{m \cdot s}{kg}$

$\rho_v := 0.1 \qquad \frac{kg}{m^3}$

$N_r := \frac{2}{3} \cdot N_r \qquad N_r = 44.149$

$$h_c := 0.95 k_L \cdot \left[\frac{\rho_L \cdot \left(\rho_L - \rho_v\right) 9.81}{\mu_L \cdot \Gamma_h}\right]^{\frac{1}{3}} \cdot N_r{}^{-\frac{1}{6}}$$

$h_c = 698.076$

Close enough to the previous value 666 W/m^2*K, so no further correction to Tw is needed.

$h_c := h_c \cdot \frac{W}{m^2 \cdot K}$ $h_c = 698.076 \frac{W}{m^2 \cdot K}$

- Calculation of Tube-side Coefficient

Tube cross sectional area $A_{cross} := \frac{\pi}{4} \cdot Di^2 \cdot Nt$ $A_{cross} = 2.713 m^2$

Density of water at $\frac{T_{tube}}{K} - 273 = 15$ °C

$\rho_1 := 998.23 \frac{kg}{m^3}$

Tube velocity $v := \frac{F_{tube}}{\rho_1 \cdot A_{cross}}$ $v = 0.106 \frac{m}{s}$

$T_t := \frac{(t_1 + t_2)}{2K} - 273$ $T_1 = 318.15K$ °C

$di := \frac{Di}{m}$

$di = 0.018$ mm

$u_t := \frac{v}{\frac{m}{s}}$ $u_t = 0.106$ m/s

$$h_i := \frac{4200(1.35 + 0.02 \cdot T_t) \cdot u_t^{0.8}}{di^{0.2}}$$

$h_i = 2.579 \times 10^3$ W/m2*K

- Overall Coefficient

$h_c = 698.076 \frac{W}{m^2 \cdot K}$

$$h_c := \frac{h_c}{\left[\frac{W}{m^2 \cdot K}\right]}$$

$U := 650$

Given

$$\frac{1}{U} = \frac{1}{h_c} + \frac{1}{6000} + \frac{\frac{Do}{m}\cdot \ln\left(\frac{Do}{Di}\right)}{2\cdot 50} + \frac{Do}{Di}\cdot\frac{1}{3000} + \frac{Do}{Di}\cdot\frac{1}{h_i}$$

$(U) := \mathrm{Find}(U)$

$U = 412.972$ W/m2*oC

Iteration is required.

$U := 40($ W/m2*oC

$Q = 2.412\times 10^{7}\,W$ $\Delta T = 24.595K$

$Q := \frac{Q}{W}$ $Q = 2.412\times 10^{7}$ $\Delta T := \frac{\Delta T}{K}$ $\Delta T = 24.595$

- Calculation of Heat Transfer Area

$A := \frac{Q}{U\cdot\Delta T}$ $A = 2.452\times 10^{3}$

- Calculation of Number of Tubes (Nt)

$A_{tube} = 0.115m^{2}$ $A_{tube} := \frac{A_{tube}}{m^{2}}$ $A_{tube} = 0.115$

$Nt := \frac{A}{A_{tube}}$ $Nt = 2.132\times 10^{4}$

- Calculation of Bundle and Shell Diameters

Constants K1 and n1:

Table 12.4 (Sinnott p.649)

$K_1 := 0.17$

$n_1 := 2.28$

$D_b := Do\cdot\left(\frac{Nt}{K_1}\right)^{\frac{1}{n_1}}$ Equation for square patterns

$D_b = 3.363\text{m}$

Use a Split-Ring Floating Head Type

From Figure 12.10, Bundle Diameter Clearance: $D_c := 150\ \text{mm}$

Shell Inside Diameter, Ds: $D_s := D_b + D_c$ $D_s = 3.513\ \text{m}$

Number of tubes in centre row:

Use square pitch $P_t := 1.25 \cdot Do$ $N_r := \dfrac{D_b}{P_t}$ $N_r = 134.538$ Tubes in centre row

$\Gamma_h := \dfrac{W_c}{L \cdot Nt}$ $\Gamma_h = 2.545 \times 10^{-4}\ \dfrac{\text{kg}}{\text{ms}}$ $\Gamma_h := \Gamma_h \cdot \dfrac{\text{m} \cdot \text{s}}{\text{kg}}$ $\Gamma_h = 2.545 \times 10^{-4}$

$\rho_v := 0.1\ \dfrac{\text{kg}}{\text{m}^3}$

$N_r := \dfrac{2}{3} \cdot N_r$ $N_r = 89.692$

$$h_c := 0.95 k_L \cdot \left[\frac{\rho_L \cdot (\rho_L - \rho_v)\, 9.81}{\mu_L \cdot \Gamma_h}\right]^{\frac{1}{3}} \cdot N_r^{-\frac{1}{6}}$$

$h_c = 781.52\ \dfrac{\text{W}}{\text{m}^2 \cdot \text{K}}$

Given

$$\frac{1}{U} = \frac{1}{h_c} + \frac{1}{6000} + \frac{\frac{Do}{m} \cdot \ln\left(\frac{Do}{Di}\right)}{2 \cdot 50} + \frac{Do}{Di} \cdot \frac{1}{3000} + \frac{Do}{Di} \cdot \frac{1}{h_i}$$

$(U) := \text{Find}\ (U)$

$U = 440.816$ W/m2*oC

Close enough to estimate firm up design.

Shell Side Pressure Drop

$\frac{T_{shell}}{K} - 273 = 41.075$

Shell inside Diameter: $D_{cl} := 150 \ mm$ Clearance

$Di_{shell} := D_b + D_{cl}$ $Di_{shell} = 3.513 \ m$

Baffle Spacing: Select baffle spacing equal to shell diameter $l_B := Di_{shell}$

Use Kern Method to make an approximate estimate: $Do = 0.02 \ m$

Cross-flow area: $A_s := \frac{(P_t - Do) \cdot Di_{shell} \cdot l_B}{P_t}$

$A_s = 2.469 m^2$

Mass Flow Rate based on the inlet conditions

$G_s := \frac{W_c}{A_s}$ $G_s = 4.023 \ \frac{kg}{m^2 s}$

$u_s := \frac{G_s}{(0.1) \frac{kg}{m^3}}$ $u_s = 40.234 \ \frac{m}{s}$

Equivalent diameter

$d_e := \frac{1.27}{Do} \cdot \left[P_t^2 - 0.785 Do^2\right]$ $d_e = 0.02m$

Vapor viscosity $\mu := 0.01$ mNs/m2

$Re := \frac{\frac{u_s}{\frac{m}{s}} \cdot \frac{d_e}{m} \cdot 0.1}{\mu \cdot 10^{-3}}$ $Re = 7.946 \times 10^3$

Friction factor will be read from Figure 12.30 (Sinnott, volume 6, third edition)

$j_f := 5 \cdot 10^{-2}$

Take viscosity drop as 50% of that calculated using inlet flow; neglect viscosity correction:

$$\Delta P_s := \frac{1}{2} \cdot \left[8 \cdot j_f \cdot \left(\frac{Di_{shell}}{d_e} \right) \cdot \left(\frac{L}{l_B} \right) \cdot \frac{\left[0.1 \cdot \left(u_s \cdot \frac{s}{m} \right) \right]^2}{2} \right] \qquad \Delta P_s = 1.5 \times 10^3 \qquad \frac{N}{m^2}$$

$\text{bar} := 10^5 \text{Pa}$

$\Delta P_s := \Delta P_s \cdot \frac{N}{m^2} \qquad \Delta P_s = 0.015\text{bar}$

Pressure of the liquid stream is:

$P_1 := 0.095 \cdot \text{bar} \qquad P_2 := P_1 - \Delta P_s \qquad P_2 = 0.08 \ \text{bar}$

Appendix 8.9 Pumps

In this section the design of pumps will be explained. There are 11 pumps in the process, the specific calculation for only one pump (P107) which pump the solution from evaporator to the crystallizer is introduced. Before the calculation of the pumps, design of pipe size is needed.

The optimum pipe diameter d_{opt} is calculacted by applying the 'economic pipe diameter' formulas.

Carbon steel pipe: $d\_optimim := 293 \cdot G^{0.53} \cdot \rho^{-0.37}$

Where G is the mass flow rate and ρ is the density of the solution

$G = 8.4 \frac{kg}{s}$

Since the concentration of the solution is low, here the density of this solution is assumed to be close to that in water.

$\rho := 1000 \frac{kg}{m^3}$

$d\_optimim = 70.264\ mm$

Real diameter for the pipe is determined according to Unit Operations, Chemical Engineering, Mccabe Smith Harriott, Appendix 3, Page 1068

Nominal pipe size:3 in. schedule no. 40, inside diameter:3.068 in.

$d := 0.08\ m$

Pipe cross-sectional area: $A := \pi \cdot \left(\frac{d}{2}\right)^2$

$\phi := \frac{G}{\rho}$

Flow velocity *u* is calculated as follow: $u := \frac{\phi}{A}$

Where ϕ is volumetric flow rate

$u = 1.671 \frac{m}{s}$

Reynold No $Re := d \cdot u \cdot \frac{\rho}{\mu}$

Where μ is the viscosity of the solution

Since the concentration of the solution is low, here the viscosity of this solution is assumed to be close to that in water.

Viscosity of water can be calculated from the equation below (Coulson and Richardson's page 947):

$$\log(\mu\_water) = VISA \cdot \left(\frac{1}{T} - \frac{1}{VISB} \right)$$

$VISA := 658.25$ $VISB := 283.16$

$T := 318.1 \ K$

$$\mu = 5.551 \times 10^{-4} \ \frac{Ns}{m^2}$$

$Re = 2.409 \times 10^{5}$

From table 5.2. (Sinnott) absolute pipe roughness is obtained, and relative roughness is calculated as follow:

$\varepsilon := 4.6 \cdot 10^{-5} \ m$

Relative roughness $\frac{\varepsilon}{d} = 5.75 \times 10^{-4}$

Pipe friction value (f) is obtained from figure 5.7.

$f := 0.002$

Pressure drop in a pipe due to friction is calculated as follow:

$$\Delta Pf := 8 \cdot f \cdot \left(\frac{L}{d} \right) \cdot \rho \cdot \frac{u^2}{2}$$

$$\Delta Pf = 9.635 \times 10^{3} \ \frac{N}{m^2}$$

Where Δp_f=Pressure drop

f = friction factor

L = pipe length including extra length of pipe to allow for miscellaneous losses

d = pipe inside diameter

ρ = fluid density

u = fluid velocity

$L := 3 \ m$

The total energy required to transport the fluid can be calculated from the following equation:

$$g\,\Delta z + \Delta P / \rho - \Delta P_f / \rho - W = 0$$

where W = work done

Δz = difference in elevations (z1-z2)

ΔP = difference in system pressures (P1-P2)

ΔP_f = pressure drop due to friction, including miscellaneous losses.

ρ = liquid density

g = acceleration due to gravity

$z1 := 3.5\,m$ $\qquad$ $z2 := 2\,m$

$$\Delta z := z1 - z2$$

$Pcrystallizer := 101325\,\frac{N}{m^2}$ $\qquad$ $P\_evaporator = 9.5 \times 10^3\,\frac{N}{m^2}$

$$\Delta P := Pcentrifuge - P\_evaporator$$

$$W := g\cdot\Delta z + \frac{\Delta P}{\rho} - \frac{\Delta Pf}{\rho}$$

$$W = -86.745\,\frac{J}{kg}$$

The pump power is calculated as follow:

$P := -W\cdot\frac{G}{\eta}$ $\qquad$ η: efficiency

Pump efficiency is estimated from Figure 5.9, in Coulson and Richardson's

$$\phi = 30.24\,\frac{m^3}{hr}$$

select pump with capacity of 50 m3/h

$$\eta := 0.7$$

$$P = 1.041 \times 10^3\,w$$

The net positive suction head can be calculated as below:

$$NPSH = \frac{P}{\rho\cdot g} + H - \frac{Pf}{\rho\cdot g} - \frac{Pv}{\rho\cdot g}$$

Where H is the height of liquid above the pump suction,

Pv is the vapor pressure of the fluid

The pressure loss in the suction piping. Only consider the friction losses in suction side

$$Pf := 8 \cdot f \cdot \left(\frac{L\_suction}{d}\right) \cdot \rho \cdot \frac{u^2}{2}$$

$L\_suction := 8\ m$

$Pf = 2.569 \times 10^3 \ \frac{N}{m^2}$

Vapor pressure correlation: Antoine-type equation from Coulson and Richardson's page 947

$$LogP = A + \frac{B}{T} + C \cdot LogT + D \cdot T + E \cdot T^2$$

$A := 29.8605$

$B := -3.1522 \cdot 10^3$

$C := -7.3037$

$D := 2.4247 \cdot 10^{-9}$

$E := 1.8090 \cdot 10^{-6}$

$T := 318.15$

$Pv = 9.597 \times 10^3 \ \frac{N}{m^2}$

$$NPSH := \frac{Pevaporator}{\rho \cdot g} + H - \frac{Pf}{(\rho \cdot g)} - \frac{Pv}{\rho \cdot g}$$

$NPSH = 3.228\ m$

APPENDIX 9 WASTES

All the waste streams in the process from the process stream summary are shown in this chapter.

Table A9.1 Waste Summary Table

Batch number per year	22						
Stream Nr. :		<116>					
Batch Cycle :		Biomass discharged			Solution discharged with biomass		
Component	Mw kg	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.
Glucose	0.180	0.0	0.000	0.000	96.7	52.237	0.237
Tyrosine	0.181	0.0	0.000	0.000	0.0	0.000	0.000
L-Phe	0.165	0.0	0.000	0.000	290.0	156.711	0.710
Acetate	0.059	0.0	0.000	0.000	42.9	23.161	0.105
Biomass	0.025	3480.7	1881.000	8.517	0.0	0.000	0.000
NH4	0.018	0.0	0.000	0.000	1.0	0.533	0.002
H2O	0.018	0.0	0.000	0.000	15421.7	8334.000	37.738
O2	0.032	0.0	0.000	0.000	0.1	0.029	0.000
CO2	0.044	0.0	0.000	0.000	0.0	0.000	0.000
N2	0.028	0.0	0.000	0.000	0.0	0.000	0.000
Total		3480.719	1881.000	0.000	15852.298	8566.671	38.791
Enthalpy :	MJ-MW-MW						
Phase :	L/V/S		S			L	
Press. :	Bara		1.0			1.0	
Temp :	oC		37			37	
Cycle times	[h]						
- Cycle & Process:			1.9	408.7		1.9	408.7

Batch number per year 22							
Stream Nr. :		<113>			<129>		
Batch Cycle :		Spent air of fermentor		Start/End	Spent hot gas off dryer		Start/End
Component	Mw kg	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.
Glucose	0.180	0.0	0.000	0.000	0.0	0.000	0.000
Tyrosine	0.181	0.0	0.000	0.000	0.0	0.000	0.000
L-Phe	0.165	0.0	0.000	0.000	0.0	0.000	0.000
Acetate	0.059	0.0	0.000	0.000	0.0	0.000	0.000
Biomass	0.025	0.0	0.000	0.000	0.0	0.000	0.000
NH4	0.018	0.0	0.000	0.000	0.0	0.000	0.000
H2O	0.018	0.0	0.000	0.000	28491.3	84.331	69.719
O2	0.032	440687.5	1233.382	1078.382	14611508.9	43248.450	35755.030
CO2	0.044	254979.9	713.630	623.947	34953.8	103.459	85.533
N2	0.028	681765.2	1908.103	1668.311	48869424.1	144648.090	119585.712
Total		1377432.590	3855.115	3370.640	63544378.099	188084.331	155495.995
Enthalpy :	MJ-MW-MW	-4541395	-3.531	-3.087	-	-	-
Phase :	L/V/S		V			V	
Press. :	Bara		1.0			1.0	
Temp :	oC		25			<125	
Cycle times	[h]						
- Cycle & Process:		-	357.3	408.7	-	337.9	408.7

Batch number per year : 22							
Stream Nr. :		<134>			<146>		
Batch Cycle :		Permeate purge stream		Start/End	Spent washing water S102 (A/B) Start/End		
Component	Mw kg	Cycle kg	Cycle kg/h	Process kg/h avg.	Cycle kg	Cycle kg/h	Process kg/h avg.
Glucose	0.180	2083.5	6.167	5.098	310.2	147.731	0.759
Tyrosine	0.181	0.0	0.000	0.000	0.0	0.000	0.000
L-Phe	0.165	5173.8	15.314	12.661	770.4	366.865	1.885
Acetate	0.059	924.0	2.735	2.261	137.5	65.453	0.336
Biomass	0.025	0.0	0.000	0.000	0.0	0.000	0.000
NH4	0.018	21.3	0.063	0.052	0.0	0.005	0.000
H2O	0.018	122024.5	361.179	298.600	30240.0	14400.000	73.999
O2	0.032	0.3	0.001	0.001	0.0	0.000	0.000
CO2	0.044	0.0	0.000	0.000	0.0	0.000	0.000
N2	0.028	0.0	0.000	0.000	0.0	0.000	0.000
Total		130227.5	385.459	318.673	31458.1	14980.054	76.979
Enthalpy :	MJ-MW-MW	-1657796	-1.363	-1.127	-410528	-54.303	-0.279
Phase :	L/V/S		L			L	
Press. :	Bara		1.0			1.0	
Temp :	oC		45			25	
Cycle times	[h]						
- Cycle & Process :		-	337.9	408.7	-	2.1	408.7

Batch number per year	22			
Stream Nr. :		140		
Batch Cycle :		Purge stream from V101		Start/End
Component	Mw kg	Cycle kg	Cycle kg/h	Process kg/h avg.
Glucose	0.180	0.0	0.000	0.000
Tyrosine	0.181	0.0	0.000	0.000
L-Phe	0.165	0.0	0.000	0.000
Acetate	0.059	0.0	0.000	0.000
Biomass	0.025	0.0	0.000	0.000
NH4	0.018	0.0	0.000	0.000
H2O	0.018	326171.3	965.431	798.156
O2	0.032	0.0	0.000	0.000
CO2	0.044	0.0	0.000	0.000
N2	0.028	0.0	0.000	0.000
Total		326171.309	965.431	798.156
Enthalpy :	MJ-MW-MW	-4364172	-3.588	-2.966
Phase :	L/V/S		L	
Press. :	Bara		1.1	
Temp :	oC		37	
Cycle times	[h]			
- Cycle & Process :		-	337.9	408.7

The vapor purge only need cooling before it goes to the environment.

APPENDIX 10 PROCESS SAFETY

Appendix 10.1 Dow F&EI Study of Dryer

Fire & Explosion Index

Area/Country: Netherlands	**Division:** -	**Location** -	**Date** 16-10-2005
Site -	**Manufacturing Unit** L-Phenylalanine Production	**Process Unit** Pneumatic Dryers	
Materials in Process Unit Water, Air,L-Phenylalanine			
State of Operation Drying the L-Phenylalanine crystal products after washing	**Basic Materials for Material Factor** L-Phenylalanine crystal dust Nf = 1 Nr = 0		
Material Factor			4

1. General Process Hazards	**Penalty Factor Range**	**Penalty Used**
Base Factor	1.00	1.00
A. Exothermic Chemical Reactions	0.30 - 1.25	0.00
B. Endothermic Processes	0.20 - 0.40	0.40
C. Material Handling and Transfer	0.25 - 1.05	0.00
D. Enclosed or Indoor Process Units	0.25 - 0.90	0.50
E. Acces	0.20 - 0.35	0.00
F. Drainage and Spill Control	0.25 - 0.50	0.00
General Process Hazards Factor (F1)		1.90

2. Special Process Hazards	**Penalty Factor Range**	**Penalty Used**
Base Factor	1.00	1.00
A. Toxic Material(s)	0.20 - 0.80	0.00
B. Sub-Atmosferic Pressure (< 500 mm Hg)	0.50	0.00
C. Operation In or Near Flammable Range		
1. Tank Farms Storage Flammable Liquids	0.50	0.00
2. Process Upset or Purge Failure	0.30	0.00
3. Always in Flammable Range	0.80	0.00

D. Dust Explosion	0.25 - 2.00	0.75
E. Pressure Operating Pressure: 100kPa Relief Setting: kPa		
F. Low Temperature	0.20 - 0.30	0.00
G. Quantity of Flammable Material: 0.2kg Hc = kcal/kg		
1. Liquids or Gases in Process		0.00
2. Liquids or Gases in Storage		0.00
3. Combustible Solids in Storage, Dust in Process		0.10
H. Corrosion and Erosion	0.10 - 0.75	0.20
I. Leakage - Joints and Packing	0.10 - 1.50	0.10
J. Use of Fired Equipment		0.40
K. Hot Oil Heat Exchange System	0.15 - 1.15	0.00
L. Rotating Equipment	0.50	0.50
Special Process Hazards Factor (F2)		3.05
Process Units Hazards Factor (F1 x F2) = F3		5.80
Fire and Explosion Index (F3 x MF = F&EI)		**23**

Appendix 10.2 Less Control Credit Factor of Dryer

Loss Control Credit Factors

Area/Country: Netherlands	**Division:** -	**Location** -	**Date** 16-10-2005
Site -	**Manufacturing Unit** L-Phenylalanine Production	**Process Unit** Pneumatic Dryers	

Materials in Process Unit

Water, Air,L-Phenylalanine

State of Operation Drying the L-Phenylalanine crystal products after washing	**Basic Materials for Material Factor** L-Phenylalanine crystal dust Nf = 1 Nr = 0

1. Process Control Credit Factor (C1)	**Credit Factor Range**	**Credit Used**
A. Emergency Power	0.98	0.98
B. Cooling	0.97 - 0.99	0.99
C. Explosion Control	0.84 - 0.98	0.84
D. Emergency Shutdown	0.96 - 0.99	0.98
E. Computer Control	0.93 - 0.99	0.99
F. Inert Gas	0.94 - 0.96	0.94
G. Operating Instructions/Procedures	0.91 - 0.99	0.94
H. Reactive Chemical Review	0.91 - 0.98	0.91
I. Other Process Hazards Analysis	0.91 - 0.98	1.00
Loss Control Credit Factor (C1)		0.63

2. Material Isolation Credit Factor (C2)	**Credit Factor Range**	**Credit Used**
A. Remote Control Valves	0.96 - 0.98	0.98
B. Dump/Blowdown	0.96 - 0.98	1.00
C. Drainage	0.91 - 0.97	1.00
D. Interlock	0.98	1.00
Material Isolation Credit Factor (C2)		0.98

3. Fire Protection Credit Factor (C3)	**Credit Factor**	**Credit**

	Range	Used
A. Leak Detection	0.94 - 0.98	0.98
B. Structural Steel	0.95 - 0.98	0.98
C. Fire Water Supply	0.94 - 0.97	0.97
D. Special Systems	0.91	0.91
E. Sprinkler Systems	0.74 - 0.97	0.88
F.Water Curtains	0.97 - 0.98	0.97
G. Foam	0.92 - 0.97	1.00
H. Hand Extinguishers/Monitors	0.93 - 0.98	0.98
I. Cable Protection	0.94 - 0.98	0.98
Loss Control Credit Factor (C3)		0.69
Loss Control Credit Factor (C1 x C2 x C3)		**0.43**

Appendix 10.3 Process Unit Risk Analysis of Dryer

Process Unit Risk Analysis			
Area/Country: Netherlands	**Division:** -	**Location** -	**Date** 16-10-2005
Site -	**Manufacturing Unit** L-Phenylalanine Production	**Process Unit** Pneumatic Dryers	
Materials in Process Unit Water, Air,L-Phenylalanine			
State of Operation Drying the L-Phenylalanine crystal products after washing		**Basic Materials for Material Factor** L-Phenylalanine crystal dust Nf = 1 Nr = 0	

1. Fire & Explosion Index (F&EI)	23	[-]
2. Radius of Exposure	6	[m]
3. Area of Exposure	111	[m^2]
4. Value of Exposure	4.00	M€
5. Damage Factor	0.10	[-]
6. Base Maximum Probable Property Damage	0.40	M€
7. Loss Control Credit Factor	0.43	[-]
8. Actual Maximum Probable Property Damage	0.17	M€
9. Maximum Probable Days Outage	9	days
10. Business Interruption	0.46	M€

Appendix 10.4 Dow F&EI Study of Vacuum Evaporator

Fire & Explosion Index

Area/Country: Netherlands	**Division:** -	**Location** -	**Date** 16-10-2005
Site -	**Manufacturing Unit** L-Phenylalanine Production		**Process Unit** Vacuum evaporator

Materials in Process Unit

Water (93.4%),L-Phenylalanine (4.4%),Glucose (16%), Acetate (0.7%)

State of Operation	**Basic Materials for Material Factor**				
Concentration solutions by vacuum evaporation	L-Phe	Nf = 1	Nr = 0	Nh = 1	MF=4
	Glucose	Nf = 0	Nr = 0	Nh = 0	MF=1
	Acetate	Nf = 2	Nr = 0	Nh = 3	MF=10
	H2O	Nf = 0	Nr = 0	Nh = 0	MF=1

Material Factor			1.20
1. General Process Hazards		**Penalty Factor Range**	**Penalty Used**
Base Factor		1.00	1.00
A. Exothermic Chemical Reactions		0.30 - 1.25	0.00
B. Endothermic Processes		0.20 - 0.40	0.20
C. Material Handling and Transfer		0.25 - 1.05	0.00
D. Enclosed or Indoor Process Units		0.25 - 0.90	0.30
E. Access		0.20 - 0.35	0.00
F. Drainage and Spill Control		0.25 - 0.50	0.00
General Process Hazards Factor (F1)			1.50
2. Special Process Hazards		**Penalty Factor Range**	**Penalty Used**
Base Factor		1.00	1.00
A. Toxic Material(s)		0.20 - 0.80	0.30
B. Sub-Atmosferic Pressure (< 500 mm Hg)		0.50	0.50
C. Operation In or Near Flammable Range			
1. Tank Farms Storage Flammable Liquids		0.50	0.00
2. Process Upset or Purge Failure		0.30	0.00
3. Always in Flammable Range		0.80	0.80
D. Dust Explosion		0.25 - 2.00	0.00
E. Pressure	Operating Pressure: 61 kPa		
	Relief Setting: kPa		0.00
F. Low Temperature		0.20 - 0.30	0.00
G. Quantity of Flammable Material:	19 kg		
Assumed	Hc = 11110 kcal/kg		
1. Liquids or Gases in Process			0.10
2. Liquids or Gases in Storage			0.00
3. Combustible Solids in Storage, Dust in Process			0.00
H. Corrosion and Erosion		0.10 - 0.75	0.20
I. Leakage - Joints and Packing		0.10 - 1.50	0.10
J. Use of Fired Equipment			0.00
K. Hot Oil Heat Exchange System		0.15 - 1.15	0.00

L. Rotating Equipment	0.50	0.30
Special Process Hazards Factor (F2)		3.30
Process Units Hazards Factor (F1 x F2) = F3		4.95
Fire and Explosion Index (F3 x MF = F&EI)		**6**

Appendix 10.5 Less Control Credit Factor of Vacuum Evaporator

Loss Control Credit Factors

Area/Country: Netherlands	**Division:** -	**Location** -	**Date** 16-10-2005
Site -	**Manufacturing Unit** L-Phenylalanine Production	**Process Unit** Vacuum evaporator	

Materials in Process Unit

Water (93.4%),L-Phenylalanine (4.4%),Glucose (16%), Acetate (0.7%)

State of Operation

Concentration solutions by vacuum evaporation

Basic Materials for Material Factor

L-Phe	Nf = 1	Nr = 0	Nh = 1	MF=4
Glucose	Nf = 0	Nr = 0	Nh = 0	MF=1
Acetate	Nf = 2	Nr = 0	Nh = 3	MF=10
H2O	Nf = 0	Nr = 0	Nh = 0	MF=1

1. Process Control Credit Factor (C1)	**Credit Factor Range**	**Credit Used**
A. Emergency Power	0.98	0.98
B. Cooling	0.97 - 0.99	0.98
C. Explosion Control	0.84 - 0.98	0.98
D. Emergency Shutdown	0.96 - 0.99	1.00
E. Computer Control	0.93 - 0.99	0.99
F. Inert Gas	0.94 - 0.96	0.94
G. Operating Instructions/Procedures	0.91 - 0.99	0.99
H. Reactive Chemical Review	0.91 - 0.98	0.93
I. Other Process Hazards Analysis	0.91 - 0.98	0.98
Loss Control Credit Factor (C1)		0.79

2. Material Isolation Credit Factor (C2)	**Credit Factor Range**	**Credit Used**
A. Remote Control Valves	0.96 - 0.98	0.98
B. Dump/Blowdown	0.96 - 0.98	0.98
C. Drainage	0.91 - 0.97	0.97
D. Interlock	0.98	
Material Isolation Credit Factor (C2)		0.93

3. Fire Protection Credit Factor (C3)	**Credit Factor Range**	**Credit Used**
A. Leak Detection	0.94 - 0.98	0.97
B. Structural Steel	0.95 - 0.98	0.87
C. Fire Water Supply	0.94 - 0.97	0.97
D. Special Systems	0.91	0.91
E. Sprinkler Systems	0.74 - 0.97	0.90
F.Water Curtains	0.97 - 0.98	0.97
G. Foam	0.92 - 0.97	0.95
H. Hand Extinguishers/Monitors	0.93 - 0.98	0.98
I. Cable Protection	0.94 - 0.98	0.98

Loss Control Credit Factor (C3)	0.59
Loss Control Credit Factor (C1 x C2 x C3)	**0.44**

Appendix 10.6 Process Unit Risk Analysis of Vacuum Evaporator

Process Unit Risk Analysis

Area/Country: Netherlands	**Division:** -	**Location** -	**Date** 16-10-2005
Site -	**Manufacturing Unit** L-Phenylalanine Production	**Process Unit** Vacuum evaporator	

Materials in Process Unit
Water (93.4%),L-Phenylalanine (4.4%),Glucose (16%), Acetate (0.7%)

State of Operation
Concentration solutions by vacuum evaporation

Basic Materials for Material Factor

L-Phe	Nf = 1	Nr = 0	Nh = 1	MF=4
Glucose	Nf = 0	Nr = 0	Nh = 0	MF=1
Acetate	Nf = 2	Nr = 0	Nh = 3	MF=10
H2O	Nf = 0	Nr = 0	Nh = 0	MF=1

1. Fire & Explosion Index (F&EI)	6	[-]
2. Radius of Exposure	2	[m]
3. Area of Exposure	7	[m^2]
4. Value of Exposure	4.00	M€
5. Damage Factor	0.08	[-]
6. Base Maximum Probable Property Damage	0.32	M€
7. Loss Control Credit Factor	0.44	[-]
8. Actual Maximum Probable Property Damage	0.14	M€
9. Maximum Probable Days Outage	8	days
10. Business Interruption	0.41	M€

Appendix 10.7 HAZOP study

Guide Word	Deviation	Possible Causes	Consequences	Action required
No	No flow	(1) Line fracture, or failure for ammonia transport	Moderate hazards (Irritation of eyes, lungs, skin)	(a) Institute regular patrolling and inspection of transfer line (b) First aid equipments are needed
		(2) Biomass recycle line 116 fracture	Pollute environment	Covered by (a)
		(3) Error in condenser E102	Not enough evaporation, pipes or joints may fracture by too much water increase in evaporator	Covered by (a), (c) Install pressure sensor or level sensor in evaporator, or install liquid overflow.
More	More flow	(4) Error in feeding 101, 103, 105, controller, Error in 133 recycle to fermentor	Overflow of Reactor R101	(d) Install level control in R-101, install liquid overflow tank.
		(5) Failure of Compressor K-101	Glucose, acetate, tyrosine, L-Phe, biomass health hazard	(e) Install filters or absorbents at the exit of the off gas.
Less	Less flow	(6) Error in heat exchanger of V101	Covered by (3)	(f) Temperature sensor in evaporator
		(7) Compressor K-101 failure	Not enough oxygen pumped in, acetate will be produced	(g) Use of flow controllers
		(8) Cake of filter is too thick, or filter is clogged	Pipes fracture, or liquid may leak from the plate and flames S102	(h) Install pressure drop sensor in the filter. Change filter or change filter cloth
	Less pressure	(9) Removal of gas or vapor by withdrawing with a blower. Fan or jet.	Vacuum collapse,	(i) relief valves and instruments (j) Designing the vessel for full vacuum conditions
		(10) Siphon of liquids	Vacuum collapse.	covered by (i)
Reverse	Reverse Flow	(11) Compressor K101 failure	Fall in line pressure, broth coming into the gas pipe	(k) Fit non-return valve in stream 110
As well as	Sparks Present	(12) Leakage in dryer	Possible dust explosion	(l) Locate far from a possible ignition source

Appendix 11 ECONOMY

Appendix 11.1 Economic Calculations

Table A.11.1: Calculation of Purchased Equipment Costs (PCE)

Equipment	Total Costs (k€)	Unit cost (k€)
Fermentor	**404.5**	
R101		404.5
Centrifuge	**213.0**	
S101		213.0
Evaporator	**260.6**	
V101		260.6
Crystallizer	**100.0**	
V102		100.0
Plate-frame filter	**67.0**	
S102*2		67.0
Dryer	**108.7**	
D101		108.7
Condensor	**237.7**	237.7
E101		
E102		
Storage tank	**34.4**	
T101		24.7
T102		8.0
T103		1.7
T104		
Pumps	**55.9**	
P101*2		2.1
P102*2		2.1
P103*2		2.1
P104*2		7.7
P105*2		7.4
P106*2		7.1
P107*2		5.8
P108*2		5.8
P109*2		5.7
P110*2		6.0
P111*2		4.2
Compressor	**148.7**	
K101*2		74.3
K102*2		74.3
Total PCE	**1630.5**	

(1) The price of the fermentor is determined according to the website:
http://www.matche.com/EquipCost/Reactor.htm

(2) The price of the centrifuge: the selected equipment is a nozzle disk centrifuge with continuous discharge of concentrated slurry. The cost of a stainless steel centrifuge is, according to table 18-152 from the Perry handbook, around 260000 US$. This price includes the drive motor. The minimum cost of installation is around 5-10% of the equipment price and 5-10%/yr of maintenance.

(3) The price of the evaporator, crystallizer, storage tank and condenser are estimated according to Coulson and Richardson's volume 6 chapter 6 and the website: http://www.matche.com.

(4) The price of the dryer is determined according to Perry handbook

(5) The price of the pumps has been calculated based with the help from http://www.mhhe.com/engcs/chemical/peters/data/ce.html

(6) The price of the compressor: Because of the less power requirements of an isothermal compressor are selected, the cooling in the stages of the compression is not taken in consideration. For the calculation of the compressor cost table 6.2 pag258 of Coulson and Richardson's is used. The selected type is a centrifugal compressor and the parameter for the cost evaluation is the power required

All the prices searched at certain year have been converted into that at year 2005 by the use of index.

Table A11.2 Calculation of Total Investment Costs

Item	Lang factor	Costs (k€)
Major equipment, total purchase cost (PCE)		**1631**
Installation	0.2	326
Process Piping	0.3	489
Instrumentation	0.2	326
Electrical	0.1	163
Buildings, process	0.1	163
Insulation	0.01	16
yard improvement	0.05	82
Auxilliary facilities	0.2	326
Total plant direct costs (TPDC)		**1891**
Design and Engineering	0.2	378
Construction	0.3	567
Total plant indirect cots (TPIC)		**946**
Total plant costs (TPC=TPDC+TPIC)		**2837**
Contractor's fee (Con1)	0.03	85
Contingency (Con2)	0.07	199
Direct fixed capital (DFC=TPC+Con1+Con2)		**3121**
Working Capital	15% of fixed capital	468
Total investment Costs (=DFC+Working Capital)		**3589**

Table A11.3 Variable Operating costs

Item	**Amount (year)**	**Unit**	**Price/Unit** **€**	**Costs (year)** **k€**
Raw materials				
Glucose[1]	7426.7	ton	245.8	1825.6
Tyrosine[2]	67.5	ton	16388.1	1105.6
Ammonia water[3]	136.0	ton	180.3	24.5
Demineralized water[4]	3895306.1	ton	0.015	57.5
Total raw materials				**3013.2**
Consumables				
Inoculation[5]	22.0	batch	138.3	3.0
Total consumables				**3.0**
Utilities				
Electricity[6]	7996800.3	MJ	0.022	176.9
Heating[7]	282130.9	ton	11.7	3305.9
Cooling[8]	8395629.7	ton	0.010	82.6
Total utilities				**3565.3**
Waste				
COD process water[9]	507.8	ton	300	**152.3**
Total waste				**152.3**
Miscellaneous operating materials				**15.6**
Shipping and packaging				**-**
Total variable costs				**6749.5**

Shipping and packaging has been set to be zero.

1. Price of glucose solution of 70% w/v is determined according to Bio separations Science and Engineering, roger G. Harrison
2. Price of tyrosine is determined according to Metabolic Engineering for Microbial Production of Aromatic Amino Acids and Derived Compounds
3. Price of ammonia is determined according to Bio separations Science and Engineering, roger G. Harrison
4. Price of demineralised water is determined according to Table 6.5 in Coulson and richardson's Volume 6, page 263
5. Price of inoculation is calculated based on the calculated amount of glucose, tyrosine and water.
6. Price of electricity is determined according to Table 6.5 in Coulson and richardson's Volume 6, page 263
7. Price of the heating steam is determined according to Table 6.5 in Coulson and richardson's Volume 6, page 263
8. Price of cooling water is determined according to Table 6.5 in Coulson and richardson's Volume 6, page 263
9. Price of ton COD per ton is selected with the help of Prof. Dr.ir. M.van Loosdrecht, TU Delft

Table A11.4 Summary of production costs

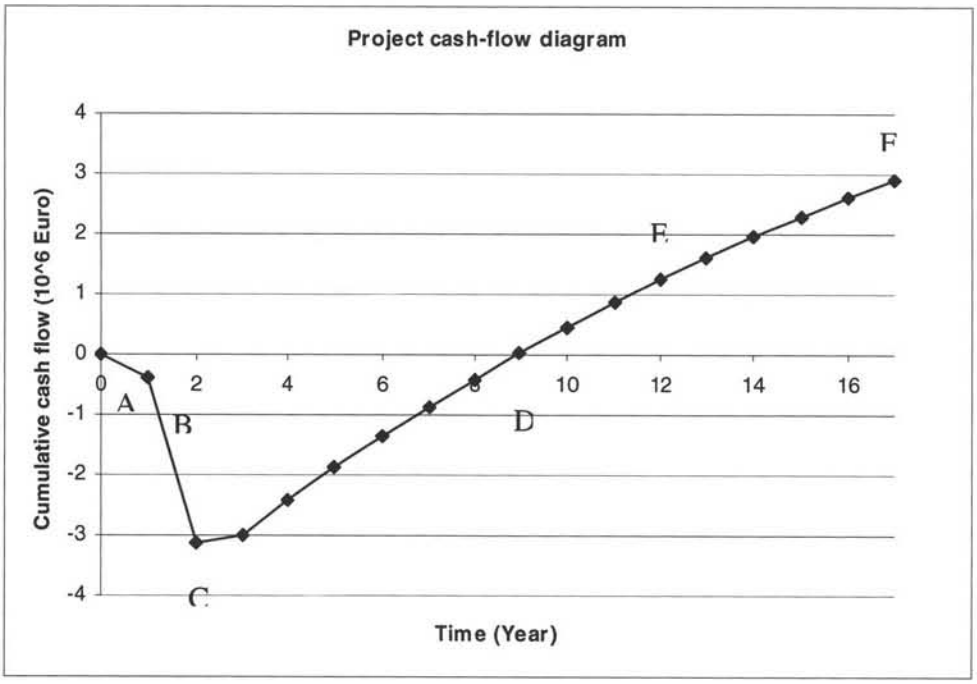
		Costs per year (k€)	Remarks
Variable costs			
Raw materials		3013.2	
Consumables		3.0	
Miscellaneous materials		15.6	10% of maintenance
Utilities		3565.3	
Waste		152.3	
Shipping and packaging		0.0	usually negligible
	Sub-total A	**6749.5**	
Fixed costs			
Maintenance		156.0	5% of the fixed capital
Operating labour		200.0	from manning estimates
Laboratary costs		40.0	20% of operating labour
Supervision		40.0	20% of operating labour
Plant overheads		100.0	50% of operating labour
Capital charges		94.6	10% of the fixed capital
Insurance		31.2	1% of the fixed capital
Local taxes		62.4	2% of the fixed capital
Royalties		31.2	1% of the fixed capital
	Sub-total B	**755.4**	
	Direct production costs	**7504.6**	9A+B
Other			
Sales expenses General overheads Research and development			Not applicable
Total production costs			
Annual (k€/a)		**7504.9**	A+B+C
Per ton Product (€/kg)		**7.5**	Annual production costs/annual production rate

Appendix 11.2 Economic Criteria

Calculation of cumulative cash flow is shown in Table A11.5. Based on these values cash flow diagram, which is shown in Figure A11.1, shows the forecast cumulative net cash flow over the life of a project. The cash flows are based on the best estimates of investment, operating costs, sales volume and sales price, which can be made for the project. Average interest rate of 5% is used to calculate the cumulative cash flow over the life of the project.

Table A11.5 Calculation of cumulative cash flow

Year (End of year)	Years Based on Year Production Start (End of Year)	Net Cash Flow NFWn (k€)	1/(1+r)^n	Net Cash Flow NPWn (k€)	Investment (k€)	Cumulative cash flow projected to NPW (M€)
1	-1	-	-	-	-378.0	-0.4
2	0	-	-	-	-2740.5	-3.1
3	1	689	0.8638	596	-467.8	-3.0
4	2	689	0.8227	567		-2.4
5	3	689	0.7835	540	-	-1.9
6	4	689	0.7462	514	-	-1.4
7	5	689	0.7107	490	-	-0.9
8	6	689	0.6768	467	-	-0.4
9	7	689	0.6446	444	-	0.0
10	8	689	0.6139	423	-	0.5
11	9	689	0.5847	403	-	0.9
12	10	689	0.5568	384	-	1.2
13	11	689	0.5303	366	-	1.6
14	12	689	0.5051	348	-	2.0
15	13	689	0.4810	332	-	2.3
16	14	689	0.4581	316	-	2.6
17	15	689	0.4363	301	-	2.9



FigureA11.1 Cumulative cash-flow diagram

A-B: The investment required to design the plant

B-C: The heavy flow of capital to build the plant, and provide funds for start-up.

C-D: The cash flow turns up at C due to the income from sales. The net cash flow is now positive but the cumulative amount still negative until the investment is paid off, at point

D: Break-even point

D-E: Cumulative net cash flow becomes positive in this region. The project is earning a return on the investment.

Calculation of DCFRR: Final interest rate of 18.4% has been determined which made the cumulative cash flow at the end of the year of the plant life to be zero.

Table A11.6 Calculation of DCFRR

Year (End of year)	Years Related to Production Start (End of Year)	Net Cash Flow NFWn (k€)	1/(1+r')^n	Net Cash Flow NPWn (k€)	Investment (k€)	Cumulative cash flow projected to NPW (M€)
1	-1	-	-	-	-378.0	-0.4
2	0	-	-	-	-2740.5	-3.1
3	1	689	0.7004	483	-467.8	-3.1
4	2	689	0.6220	429		-2.7
5	3	689	0.5523	381	-	-2.3
6	4	689	0.4905	338	-	-2.0
7	5	689	0.4356	300	-	-1.7
8	6	689	0.3868	267	-	-1.4
9	7	689	0.3435	237	-	-1.2
10	8	689	0.3051	210	-	-0.9
11	9	689	0.2709	187	-	-0.8
12	10	689	0.2406	166	-	-0.6
13	11	689	0.2137	147	-	-0.4
14	12	689	0.1898	131	-	-0.3
15	13	689	0.1685	116	-	-0.2
16	14	689	0.1497	103	-	-0.1
17	15	689	0.1329	92	-	0.0

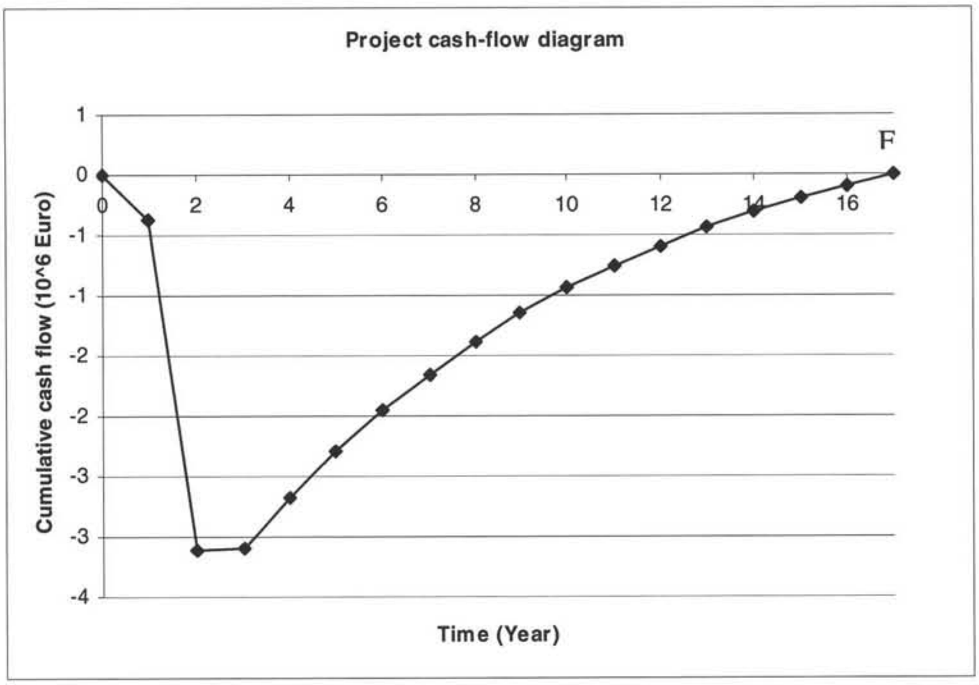


Figure A11.2 Cumulative cash-flow diagram for the calculation of DCFRR

APPENDIX 12 ALTERNATIVE PROCESSES

Appendix 12.1 Microfiltration

Micro filtration for biomass circulation

Cross flow hollow fiber cartridges are chosen as the cell separation methods in this process. Centrifugation may be used for such separations; however, additional filtration steps may still be needed before subsequent process

According to: Clarification of Pichia pastoris using cross flow micro filtration, GE healthcare. Membrane resistance (R_m) of 0.2 μm filter can be estimated as follows:

$\Delta P_{tmp1} := 2.5 psi$ Trans membrane pressure.

$\Delta P1 := \Delta P_{tmp1} + 1 \cdot 10^5 Pa$ psi is stand for psig here

$\Delta P_{tmp2} := 10 psi$

$\Delta P2 := \Delta P_{tmp2} + 1 \cdot 10^5 Pa$

Water flux:

$$J_{w1} := \Delta P_{tmp1} \cdot 70 \frac{L}{m^2 hr \cdot psi}$$

$$J_{w2} := \Delta P_{tmp2} \cdot 55 \frac{L}{m^2 hr \cdot psi}$$

$\mu_w := 0.8937 \cdot 0.01 poise$ Viscosity of water at 25°C

$$R_{m1} := \frac{\Delta P1}{J_{w1} \cdot \mu_w}$$

$$R_{m1} := \frac{\Delta P1}{J_{w1} \cdot \mu_w}$$

Assume the average value of Rm1 and R_m is the resistance of the 0.2μm membrane, therefore:

$R_m := \frac{(R_{m1} + R_{m2})}{2}$ $R_m = 1.968 \times 10^{12} m^{-1}$

CFP-2-E-154M is chosen, because it is the maximum size in the hollow fiber cartridges for membrane separation. Pore size: 0.2μm

$d_f := 1mm$ Fiber ID: 1 mm

$L_f := 110cm$ Nominal flow path length

$Am := 19m^2$ Area of one fiber, [m^2]

$\phi mE := 5.2233\,10^4 \frac{kg}{hr}$ Required permeate mass flow rate

$\rho_1 := 1000 \frac{kg}{m^3}$ The density of the broth is assumed the same as water

$\phi_{per\_total} := \frac{\phi mE}{\rho_1}$

Assume 10 cartridges are used

$n := 1($

$A_{m\_need} := n \cdot Am$ Total area of the cartridges

$J_{per} := \frac{\phi_{per\_total}}{n \cdot Am}$ Permeate flux through the micro filter

$J_{per} = 274.911 \frac{L}{m^2 hr}$

Four nominal feed stream flow rates (liters/minute) are shown below:

$\phi_{feed1} := 70 \frac{L}{min}$ feed stream flow rate, shear rate ~ 2,000 sec^-1

$\phi_{feed2} := 140 \frac{L}{min}$ feed stream flow rate, shear rate ~ 4,000 sec^-1

$\phi_{feed3} := 280 \frac{L}{min}$ feed stream flow rate, shear rate ~ 8,000 sec^-1

$\phi_{feed4} := 560 \frac{L}{min}$ feed stream flow rate, shear rate ~ 16,000 sec^-1

Estimate residence time in the filter

$A_f := \pi \cdot d_f \cdot L_f$ Area of one fiber, [m^2]

$n_f := \frac{Am}{A_f}$ Number of fiber in one cartridge

$V_f := \pi \cdot \left(\frac{d_f}{2}\right)^2 \cdot L_f$ Volume of one fiber [m^3]

$V := V_f \cdot n_f$ Total hollow fiber volume in one

	cartridge [m^3]
$\tau 3 := \frac{V}{\phi_{feed3}}$	Residence time if Φfeed3 is used
$C_{o2\_fermentor} := 0.103 \frac{mol}{m^3}$	Dissolved oxygen in the fermentor
$OUR := 130 \frac{mol}{m^3 \cdot hr}$	Oxygen uptake rate
$C_{o2\_consumed} := OUR \cdot \tau$	Oxygen consumed in the fiber
$C_{o2\_left} := C_{o2\_fermentor} - C_{o2\_consumed}$	Dissolve oxygen concentration in the exit of the cartridge
$C_{o2\_left} = 0.066 mol m^{-3}$	In order to prevent oxygen limitation during the filtration (acetate will be produced in oxygen limitation), residence time should be as shorter as possible. However the higher shear rate due to this higher flow rate may have bad effects on the cells. So flow rate 2 and 3 may be the better choice in our case.

Assume the biomass concentration in the outflow is 50 gm/L

$C_{rent\_x} := 50 \frac{gm}{L}$	Biomass concentration in the exit of the cartridge
$C_{feed\_x} := 30 \frac{gm}{L}$	Biomass concentration in the inflow
$\phi_{feed} := \phi_{feed3}$	Volume flow rate of the inflow
$\phi_{per} := \frac{\phi_{per\_total}}{n}$	Permeate flow rate in one cartridge
$C_{rent\_x} \cdot (\phi_{feed} - \phi_{per}) = \phi_{feed} \cdot C_{feed\_x}$	
$\phi_{feed} \cdot n = 130.582 \frac{m^3}{hr}$	Total incoming flow rate if reach the required permeate rate
$\frac{\phi_{per}}{\phi_{feed}} = 0.4$	

$\mu_{per} := 0.653\,0.01\text{poise}$ Viscosity of permeate, use viscosity of 40oC water, assume this is the same viscosity of the 37oC broth permeate

$\alpha := 10^{13}\,\frac{m}{kg}$ Specific cake resistance. α has a large value (10^12~10^15). A middle value is chosen.

$w := 10\cdot10^{-6}\cdot1000\frac{kg}{m^2}$ Mass cake per unit area. Assume thickness of the deposited layer is 10μm, and the density is as pure water.

$R_c := \alpha\cdot w$ Resistance of deposited layers [m^-1]

According to Darcy's law, flow through porous media

$\Delta P := \mu_{per}\cdot\left(R_c + R_m\right)\cdot J_{per}$

$\Delta P = 1.018\text{atm}$ Tran membrane pressure

Maximum transmembrane pressure in ProCell ultra filtration cartridge: 4.4 bar

Capital cost estimate micro filters

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Purchase cost of micro filtration hardware. (1998 price)

Each cartridge house:

$Cost_h := 2500($ €

$Cost_{h\_total} := Cost_h\cdot n\cdot 2$ Two times more cartridges are needed for cleaning and changing in the continuous process

$Cost_{h\_total} = 5\times10^5$ € (1998 price)

$Cost_{h\_total\_2005} = 5.969\times10^5$ € (2005)

Appendix 12.2. Hydro-Cyclone Rotary Drum Filter

In order to evaluate the alternative of the use of a Hydrocyclone to concentrate the L-Phenylalanine crystals we proceed as follows:
Using the Nomogram in Coulson and Richardson's page 421 we assume a Diameter of the particle for which the cyclone is 50% efficient

$\rho cr := 1290 \frac{kg}{m^3}$

$\rho 1 := 1000 \frac{kg}{m^3}$

$d_{50} := 25 \times 10^{-6} m$

$\rho cr - \rho 1 = 0.29 \frac{gm}{cm^3}$ Density difference between the solid and the liquid

$\phi mv := 28.2 \cdot \frac{m^3}{hr}$

$\phi mv = 470 \frac{L}{min}$ Volume flow

$\mu := 0.8937 10^{-3} \cdot \frac{N \cdot s}{m^2}$ Water viscosity

$D_h := 25 \cdot cm$ Diameter of the hydrocyclone calculated using the nomogram

The liquid leaving the hydrocyclone is assumed to contain 18-20% solids

Mass balance around the hydrocyclone:

$\phi mc := 135 \cdot \frac{kg}{hr}$

$X_{in} := \frac{\phi mc}{\phi mv \cdot \rho 1}$

$X_{in} = 0.479\%$ Inlet fraction of crystals in the feed of the hydro-cyclone

$\phi m := \phi mv \cdot \rho l$ $X_e := 0$

$\phi m = 7.833 \frac{kg}{s}$

$X_c := 18\%$ Fraction of crystals leaving the hydro-cyclone

$\phi 2 := 2 \cdot \frac{kg}{hr}$ $\phi w := 3 \cdot \frac{kg}{hr}$

Given

$\phi m = \phi w + \phi 2$

$\phi m \cdot X_{in} = \phi w \cdot X_e + \phi 2 \cdot X_c$

$$\begin{bmatrix} \phi w \\ \phi 2 \end{bmatrix} := \mathrm{Find}\begin{bmatrix} \phi w \\ \phi 2 \end{bmatrix}$$

$$\begin{bmatrix} \phi w \\ \phi 2 \end{bmatrix} = \begin{bmatrix} 2.745 \times 10^4 \\ 750 \end{bmatrix} \frac{kg}{hr}$$

Rotary drum filter:

Using a rotary drum with the following characteristics:

The resistance of the filter is neglected compared to the resistance of the cake
The pressure drop is constant

$f_f := 30\%$ Mass Fraction of the filter submerged

$f_{Phe} := 60\%$ Mass fraction of solids in the cake

$\Delta P := 0.7 \cdot atm$ Pressure drop

$\phi mc := 135 \cdot \frac{kg}{hr}$

$\phi mc\_w := 225 \cdot \frac{kg}{hr}$

$\phi 2 = 750 \frac{kg}{hr}$ Total mass flow to filter

$\rho cr := 1.29 \cdot 10^{3} \cdot \frac{kg}{m^{3}}$

$\rho l := 1000 \frac{kg}{m^{3}}$

$\rho s2 := \left[\frac{\phi mc}{\phi 2} \cdot \rho cr + \frac{\phi 2 - \phi mc}{\phi 2} \rho l\right]$

$\rho s2 = 1.052 \times 10^{3} \frac{kg}{m^{3}}$ Mass flow from the fermentor to the filter

The viscosity of the slurry will be assumed to be equal to water

$\mu := \left(0.8937 \cdot 10^{-3}\right) \cdot Pa \cdot s$ $Pa \cdot s$

The volume flow of the stream is:

$\phi v\_f2 := \frac{\phi 2}{\rho s2}$

$\phi v\_f2 = 0.713 \frac{m^{3}}{hr}$

In order to fulfill the requirement of 200-500 L/m2*hr we divide the stream in to n flows

$n := 1$

$\phi m\_fi := \frac{\phi 2}{n}$

$\phi m\_fi = 750 \frac{kg}{hr}$

$\phi v\_fi := \frac{\phi m\_fi}{\rho s2}$

$\phi v\_fi = 712.792 \frac{L}{hr}$

To calculate the concentration of crystals in the slurry we proceed:

$f_{Phe\_s} := \frac{\phi mc}{\phi m\_fi}$

$f_{Phe\_s} = 18\%$ Mass Fraction of the crystals in the slurry

$r := \frac{\phi mc\_w}{\phi mc}$ Ratio between the wet cake and the dry cake

$r = 1.667$

To calcutate the area needed to filtrate ϕv_f2 of slurry:

$s := 0$ Compressibility constant

$\alpha 1 := 1 \cdot 10^{11} \cdot \frac{m}{kg}$ Empirical constant for the determination of the cake resistance

$\alpha := \alpha 1 \cdot \left(\frac{\Delta P}{Pa} \right)^{s}$

$\alpha = 1 \times 10^{11} \frac{m}{kg}$ Specific cake Resistance

To calculate the flowrate of the filtrate we proceed as follows:

$Cs := \frac{\phi mc}{\phi v\_fi}$ kg of crystals/m3 of filtrate

$Cs = 189.396 \frac{kg}{m^3}$

$\phi f := \phi m\_fi \cdot \frac{f_{Phe\_s}}{Cs}$ Flow of filtrate of filter

$\phi f = 0.713 \frac{m^3}{hr}$

Using eq. 14.2-24 Geankoplis:

According to handouts of Bioseparations course the flux is around 200-500 L/m2*h

$t := 35\cdot hr$

Varying time

$J1 := 250\frac{L}{m^2\cdot hr}$ $\phi f = 712.792\frac{L}{hr}$

$A := \frac{\phi f}{J1}$

$A = 2.851m^2$

Given

$$A = \frac{\phi f}{\left(\frac{2\cdot f_f\cdot(\Delta P)}{t\cdot\mu\cdot\alpha\cdot Cs}\right)^{0.5}}$$

$t := Find(t)$

$t = 8.689min$ Total cycle time

$tf := f_f\cdot t$

total cycle time

$tf := f_f\cdot t$

$tf = 156.402s$ Time for formation of the cake

$$d := \frac{\frac{\phi mc}{n}\cdot t}{\rho cr\cdot A}$$

$d = 5.315mm$