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Contribution to the study of biological products extraction of industrial interest

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Lo semejante disuelve a lo semejante

Heurístico popular en química

Gracias a Jordi y Alexandra por su apoyo, críticas constructivas y por marcarme el camino hacia donde puedo seguir aprendiendo.

Gracias a mi familia por su apoyo constante.

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SUMMARY

Vanillin is an important building block for the chemical, pharmaceutical and food industries. It has the potential to become a key intermediate compound for the synthesis of bio-based polymers. Nowadays vanillin is mostly produced from petro-based resources as raw materials. This is not a sustainable situation, though, because of the nonrenewable nature of the resources employed. Therefore it would be desirable to produce it on an industrial scale from biomass (lignins). Currently, only 15% of vanillin is produced in this way, while the majority of lignins are usually burned for energy in the pulping process industry. Thus, the need for alternative solutions coming from the processing of the biomass feedstock is both a technological and economical priority. The petrochemical process is a mature process while the lignin process has yet to be improved before becoming a competitive one able to substitute the petrochemical route.

The depolymerization of Kraft lignins is being heavily studied as a possible feedstock, following the biorefinery approach that considers lignins as high added value aromatic building blocks instead of being used as fuel for energy generation.

The present work studies the efficiency of a standard Kraft lignin-derived vanilllin extraction process that includes a liquid-liquid extraction unit operation. It also proposes three process design alternatives, one exclusive to the use of aliphatic alcohols C_6-C_8 as solvents, which have physical properties that can help to further simplify the process design. Simulations are performed in AspenPlus® V10 using both simplified and rigorous mathematical models. A solvent screening process for the liquid-liquid extraction step is also performed, following a literature research on suitable solvents both industrially used and theoretically suggested. Both resource requirements and toxicity related metrics are employed for the final solvent viability classification proposed.

Keywords: Vanillin, Kraft Lignin, Biological Process Design and Optimization, Solvent Screening for Liquid-Liquid Extraction.

RESUMEN

La vanillina es un importante compuesto tanto para la industria química como para la farmacéutica y alimentaria. Asimismo, tiene un alto potencial para convertirse en un compuesto intermedio clave para la síntesis de biopolímeros.

Hoy en día, la vanillina se produce sobre todo a partir de sustratos petroquímicos. Esto no es una situación deseable ni sostenible, ya que es una fuente de recursos finita y no renovable. También puede ser producida a escala industrial a partir de biomasa (lignina). Actualmente, tan sólo un 15% de la vanillina producida proviene de dicha vía. La despolimerización de las ligninas provenientes del proceso Kraft se está estudiando intensamente como posible alternativa, valorando la lignina como un compuesto aromático de alto valor añadido en lugar de combustible para generación energética.

El presente trabajo estudia la eficiencia de un proceso de producción estándar de vanillina a través de lignina Kraft. El proceso incorpora una etapa crucial de extracción líquido-líquido. También se proponen tres alternativas de diseño, una exclusiva para el uso de alcoholes alifáticos C_6-C_8 como solventes. Se realizan simulaciones en AspenPlus® V10 utilizando modelos matemáticos tanto simplificados como rigurosos. También se realiza un proceso de cribado de solventes para la etapa de extracción líquido-líquido, siguiendo los resultados bibliográficos hallados en cuanto a solventes utilizados tanto en la práctica industrial como sugeridos teóricamente.

En éste proceso se tienen en cuenta tanto los requerimientos materiales (caudal mínimo de solvente necesario, solubilidad del solvente en el refinado) como las propiedades toxicológicas (toxicidad acuosa, LC50, parámetro de Impacto Ambiental Potencial (PEI)) de los solventes para proponer una clasificación final sobre la viabilidad de los mismos como agentes extractores de vanillina en un proceso comercial.

Palabras clave: Vanillina, Lignina Kraft, Diseño y Optimización de Procesos Biológicos, Cribado de Solventes para Extracción Líquido-Líquido.

1. INTRODUCTION

Vanillin is a chemical compound used as a building block in a variety of industries, including the chemical, pharmaceutical and food industries. At the same time, it is currently one of the only molecular phenolic compounds produced on an industrial scale from biomass. It also functions as an aromatic monomer for various synthesis processes, which is beneficial for reaching the thermo-mechanical properties desired. Because all of this, it has the potential to become a key intermediate compound for the synthesis of bio-based polymers *(Figure 1).*

Petro-based resources are the current raw materials of choice for the production of organic chemicals and polymers such as vanillin. This situation, though, is not a sustainable one because of the nonrenewable nature of the resources employed. The petroleum available will unavoidable decrease with time and this will incite an ever-increasing demand that coupled with price increases and supply problems will lead to unpredictable repercussions for the chemical industry and society as a whole. Thus, the need for alternative solutions coming from the processing of the biomass feedstock is both a technological and economical priority.

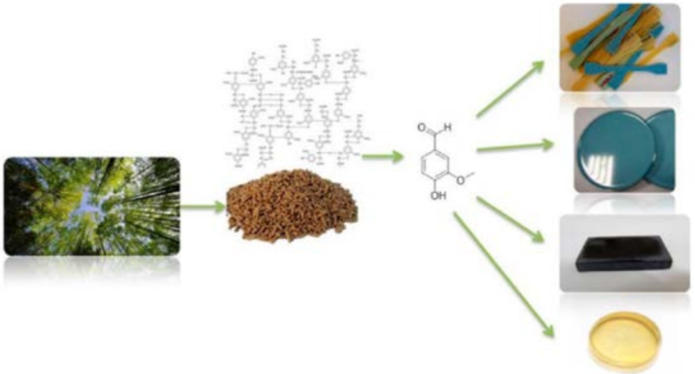


Figure 1. Vanillin sourcing and products derived from it *(Fache et al., 2016)*.

Similar to the biorefinery concept, the idea is to turn (bio)chemically each component of the feedstock into a variety of useful products. These bio-based chemicals are already industrially available, but they still face a series of production limitations that must be solved before they become well stablished processes and see their full potential developed.

Most of currently available bio-based compounds are aliphatic or cycloaliphatic, derived commonly from cellulose, starch or triglycerides. Many key chemicals are, however, aromatic and still ultimately derived from petroleum. Also, aromatics have experienced a price increase in recent years coming from a reduced availability due to yield decrease in petroleum processing processes. In this context, the need for finding biomass-derived aromatic building blocks is exacerbated.

There are three major classes of phenolics from renewable resources: lignin, tannins and cashew nutshell liquid (CNSL) *(Figure 2)*. The first two are extracted from wood. The highest commercial availability and potential is for lignin.

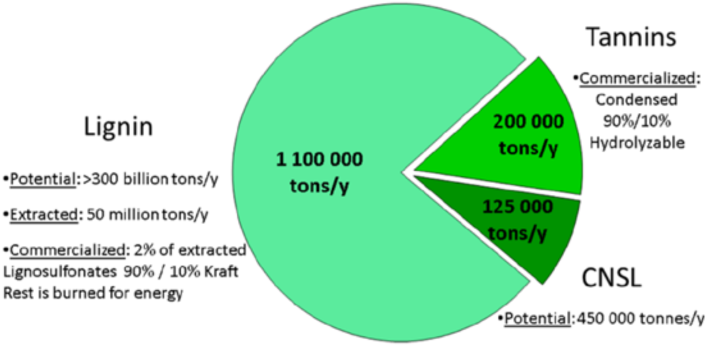


Figure 2. Commercial availability of lignin, tannins and CNSL *(Fache et al., 2016). Note: y:year.*

It is evident at first sight the discrepancy between the enormous amount of (poly)phenolic materials available from biomass and their relatively underdeveloped industrial use compared to aliphatic resources like vegetal oil. This is especially clear in the case of lignin and tannins.

It is obvious that they must have some kind of drawbacks that prevent their widespread adoption. For one, CNSL annual production volume might be insufficient to make it a reliable renewable source. This is not the case, however, for lignin and tannins. Their main problems come from their very complexity as chemical compounds, as well as their processing difficulty and purity variability depending on the plant and method of extraction. Thus, guaranteeing product consistency in time might be difficult.

Until solutions for these problems are found, for the time being, a more viable solution seems to be the handling of molecular compounds isolated from the depolymerization of these feedstocks. This is especially true in the case of the depolymerization of lignin, which is being presently heavily studied *(Mota et al., 2016)*.

Vanillin is the most available pure monoaromatic phenol that is currently produced from lignin depolymerization at industrial scale. Around 20.000 tons/year of vanillin are produced, with 15% of it coming from lignin *(Fache et al., 2016)*. The advantages of vanillin as an aromatic building block include that it is a safe compound that bears two reactive functions that can be chemically modified. Considered as a difunctional compound it is useful to prepare thermoplastic polymers. It is also, as noted earlier, a renewable resource that comes already produced in lignin and doesn’t compete with food sources.

1.1. VANILLIN PRODUCTION FROM LIGNIN

The lignin-to-vanillin depolymerization process represents, as mentioned above, 15% of the overall vanillin production. More specifically, it comes from its lignosulfonates.

Industrially, only lignin from the sulfite pulping process is used for vanillin production, despite the fact that this kind of lignin accounts for only around 10% of total lignins extracted *(Fache et al., 2016).* Kraft lignins, which are the vast majority of lignins, are usually burned for energy in the pulping process. An ever increasing amount of research is, however, investigating the depolymerization of all kinds of lignins, such as the ones coming from the Kraft process. This is consistent with the biorefinery approach that considers lignins as a source of medium-to-high added value aromatic building blocks instead of fuel for energy generation.

Consistent with this aim, this work uses lignins and the major components derived from the Kraft process, obtained under alkaline and oxidative conditions, as the source from which vanillin is extracted.

With this in mind, the first step in the process for vanillin production consists of treating an aqueous solution of lignin with oxidants at alkaline pH, at high temperature and pressure. The base lignin is thus depolymerized, yielding a complex and heterogeneous mix of lignins. This complexity makes the establishment of a reaction mechanism very challenging. The mechanism shown in *Figure 3* has been proposed in the literature, but it should be stated clearly that it is still a widely debated and researched, not completely understood, subject.

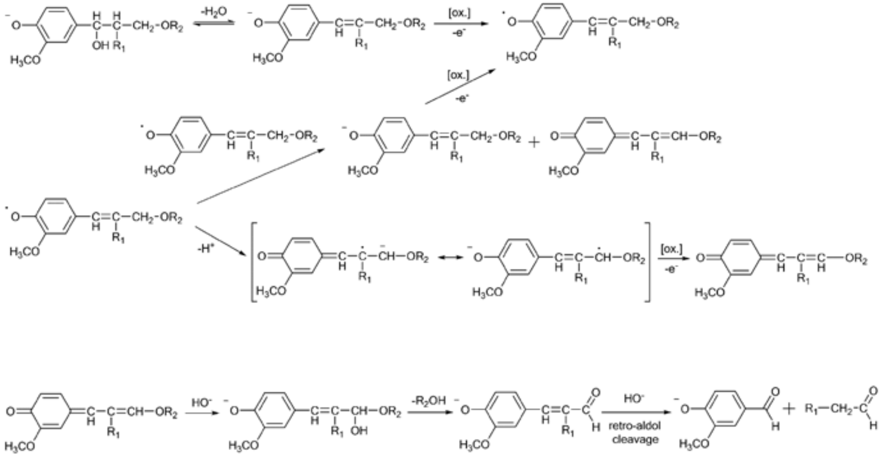


Figure 3. Proposed mechanism of lignin depolymerization leading to vanillin *(Fache et al., 2016)*.

The main parameters influencing this alkaline oxidative process have been reviewed elsewhere *(Fache et al., 2016; Mota et al., 2016).* Parameters related to chemical engineering considerations, such as batch / continuous operation mode and shape / type of reactor also play a major role in the final yield of the reaction. Although most of experimental evidence in literature has been produced on batch reactors *(Araújo et al., 2010)*, from an industrial point of view the study of continuous modes of operation is more attractive. This is due to the large liquor volumes that are treated in the pulp industry from which the Kraft lignin comes from and the easier control and homogeneous final product characteristics that are obtained operating in continuous mode. It also implies a lower overall economic investment and operating costs.

Given the complexity of the produced mixture, the downstream treatment of the alkaline oxidative lignin depolimerization reaction is by itself a complex topic that is still extensively discussed both in patents and academic literature (e.g. Silva et al., 2010; Wongtanyawat et al, 2018).

1.1.1. Downstream treatment

Appendix 1 provides a list of the most common products of alkaline oxidative lignin depolimerization usually studied in the cases of both softwoods and hardwoods. It is a complex mixture containing chemical compounds of diverse physico-chemical natures.

A way for the isolation of each of these compounds is not currently possible nor economically viable. Thus, the efforts have been mainly directed toward the extraction of the more interesting subproducts: vanillin and syringaldehyde *(Silva et al., 2010)*. We can infer the difficulty of isolating pure vanillin by the number and variety of separation processes proposed in the literature *(Mota et al., 2016)*. The two major problems are the acidification of the mixture and the residual lignin removal.

One of the oldest separation methods involves lignin precipitation in acidic conditions and then phenolics extraction by organic solvents *(Figure 4)*.

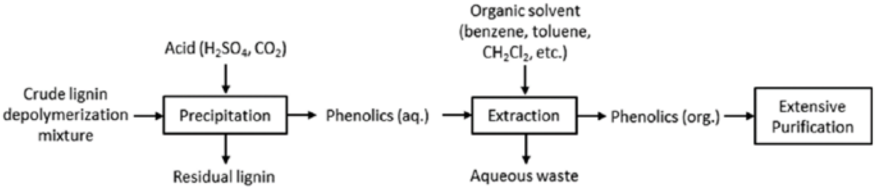


Figure 4. Lignin-to-vanillin process involving acidification and an extraction step by organic solvents *(based on Fache et al., 2016).*

Another method dating from the same time involves the bisulfitation of the mixture. Briefly explained, it consists in mixing the crude lignin depolymerization mixture with a solution of $NaHSO_3$ (sodium bisulfite) to prepare a *"vanillin bisulfite complex"*. The derivatives produced have good solubility in water, as opposed to the other products present in the mixture. Products with high molecular weight precipitate and are removed. The hydrosulfite anion reacts selectively with the aldehyde moiety, which causes vanillin and syringaldehyde to be both extracted. Finally, an acidification of the aqueous phase takes place to recover aldehydes and SO_2 *(Figure 5)*.

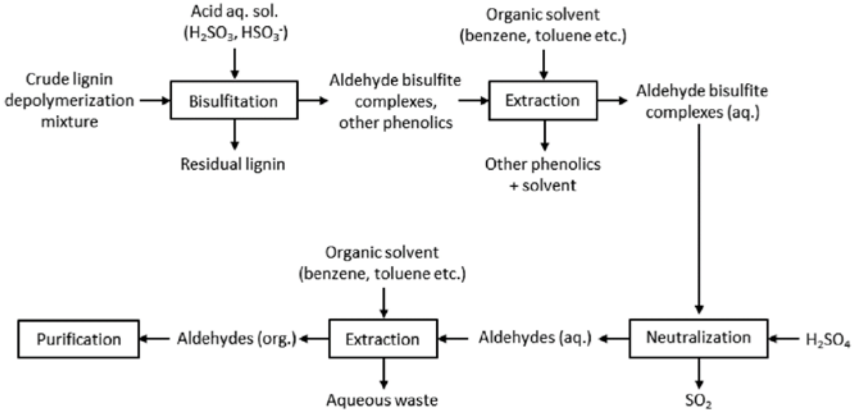


Figure 5. Lignin to vanillin process involving bisulfitation and an extraction step by organic solvents *(based on Fache et al., 2016).*

One of the biggest drawbacks of these two methods is that they use large amounts of organic solvents. Because of this, some methods that avoid this problem have been proposed. The two most commonly studied are the ones based on the use on supercritical CO_2 *(Figure 6)* and zeolites / macroporous resins (*Figure 7*) instead of organic solvents.

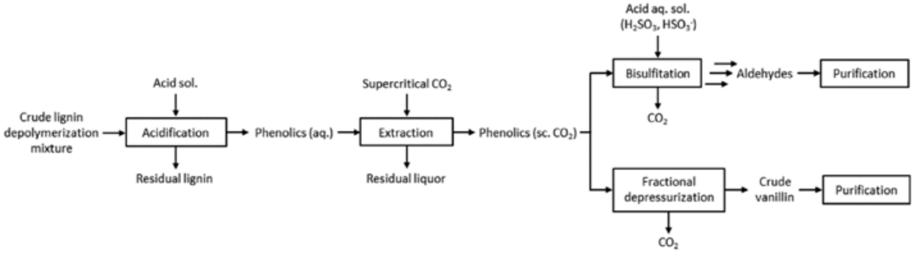


Figure 6. Lignin-to-vanillin process involving an extraction step by supercritical CO_2 *(based on Fache et al., 2016).*

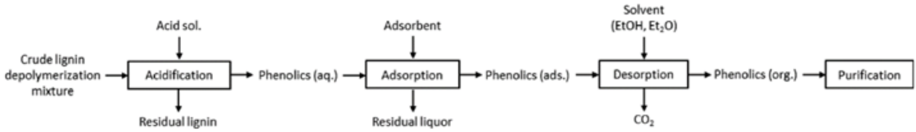


Figure 7. Lignin-to-vanillin process involving an extraction step by adsorption *(based on Fache et al., 2016).*

The major disadvantage of these alternatives are the requirement of large amounts of acids for neutralization / acidification prior to the extraction of vanillin.

A review of the methods are provided by *Fache et al. (2016) and Mota et al. (2016).* There is not a single easy solution that fixes all the problems at the same time without creating some new ones of its own.

The product stream coming out from all these methods is composed of crude vanillin with varying amounts of other phenolics, depending on which process is used. Further purification of vanillin is required to achieve technical or food purity grades, and this may be a difficult task. This is because the other components present have very close structures and properties to the vanillin ones. More information on the difficulties and solutions proposed is available by *Mota et al. (2016*). As an example, multistage crystallization and a vanillin-molecularly imprinted polymer adsorbent have been proposed *by Žabková, et al. (2007) and Zhao et al. (2012).*

1.1.1.1. Solvent selection for the extraction step

From an industrial point of view, the use of Kraft lignin as a lignin source is attractive because more than 90% of the total lignins produced worldwide come from this route (*Pinto et al., 2012*). The downstream separation in such a process normally uses ethyl acetate as solvent because it shows a high percentage recovery of vanillin product. It has the disadvantage, however, that a large amount of water is required to remove methanol completely from the products prior to the extraction step. To circumvent this disadvantage and enhance the overall process several improved processes have been proposed in the literature (e.g. *Silva et al., 2009*).

Some of these proposals involve process intensification. This approach focuses on the redesign of existing processes to allow for a better exploitation of the input raw materials, lower energy consumption and a reduction in the plant size. The objective is to create a new process based on an existing one that is more compact, more energy efficient and cleaner.

Wongtanyawat et al. (2018) applied the process intensification concept to a conventional vanillin production scheme and proposed three alternatives that were compared between them and the original process. Their modifications were centered on the extraction step and the subsequent solvent recovery through a distillation column.

Their first proposal was the base-case design, which uses ethyl acetate as extractive agent and is used as a starting point for subsequent modifications *(Figure 8).*

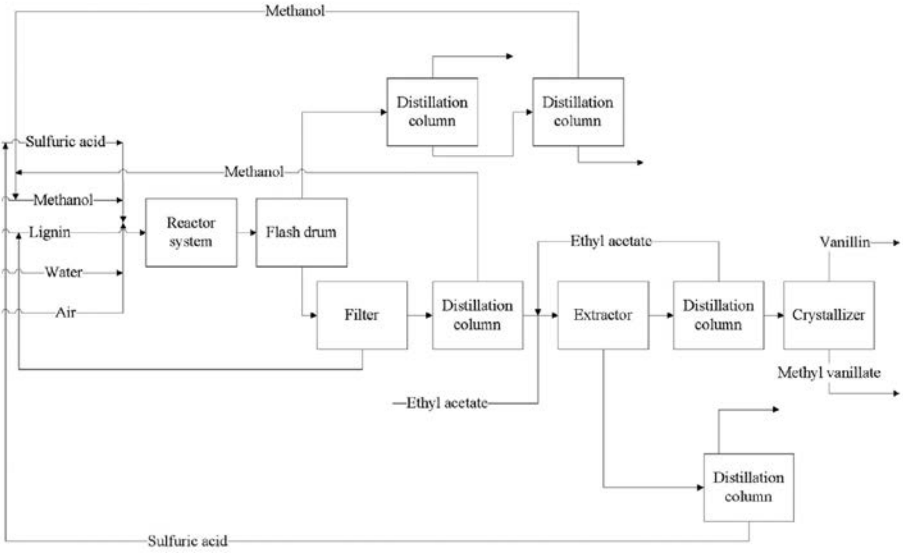


Figure 8. Block flow diagram of a vanillin production process from Kraft lignin using ethyl acetate as solvent (*Wongtanyawat et al., 2018)*.

The flash drum and the first distillation column are used to separate the gases generated in the reaction from the stream that contains the vanillin.

The second distillation column recovers methanol separating it from water. The bottoms stream of the flash unit passes through a filter that recovers the unreacted lignin that hasn't been depolymerized in the reactor.

Then, a second column for the recovery of methanol follows before the stream enters the liquid-liquid extraction unit. As mentioned, ethyl acetate is used as solvent, and then recovered in the column that receives the extract stream from the extractor unit.

The raffinate stream is sent to another column that recovers the sulfuric acid and recirculates it. As a final purification step they utilized a crystallizer unit operation to separate vanillin (principal product) and methyl vanillate (valorized subproduct) from the extract stream once the solvent has been recovered.

An alternative proposed by *Wongtanyawat et al. (2018)* only involved the substitution of the ethyl acetate solvent by benzene, but this chemical has numerous disadvantages such as toxicity. The processes were rigorously simulated in AspenPlus and both energy efficiency and environmental impact metrics were used to assess their performance. Energy consumption between the two processes was found to be very close but, even if benzene is a commercially viable option, it presents both toxicity and phase-separation difficulties (*Kaygorodov et al., 2010)*. The use of other solvents for extraction was not explored in this work.

Kaygorodov et al., 2010, however, reviewed the advantages and disadvantages of known vanillin extraction solvents *(Table 1)* and proposed the used of aliphatic alcohols C_6-C_8 over benzene and other common extractants.

They affirm that these alcohols present substantial advantages over common solvents and that they can be used in weakly alkaline media. This last point is important, because it implies that reaction solution streams of lignin oxidation can be used without their acidification.

This would eliminate the technological problems related to lignoacids precipitation that follows acidification and has the potential to reduce energy consumption greatly, as fewer columns would be needed.

Another advantage of the use of these alcohols would be that they are practically insoluble in water, which is good as the solvent lost in the raffinate from the extraction column is a metric for the solvent environmental impact.

Table 1. Characteristics of methods of vanillin extraction *(Kaygorodov et al., 2010).*

Extractant	Distribution Coefficient (D)	Advantages	Disadvantages
Benzene	*6.3*	*Industrially used*	*Toxicity of extractant, phase-separation difficulties*
Toluene	*4.1*	*Moderate toxicity*	*Low D*
Hexane	*0.2*	*High extraction selectivity*	*Very low D*
Chloroform	*26.5*	*High D value,* *quick phase separation*	*Low extraction selectivity*
Butyl acetate	*28.2*	*High D value*	*Low extraction selectivity,* *stripping difficulties*
Butanol	*27*	*High D value*	*Solubility of extractant in water*
Octanol	*20.5*	*High D value*	*Low extraction selectivity,* *stripping difficulties*
Octylamine	*Up to 600*	*High extraction selectivity and distribution ratio*	*Stripping difficulties*

2. OBJECTIVES

The present project has the following principal aims and objectives:

- To search for an existing bioprocess with a renewable source material and try to design a viable alternative that improves it in some aspect. Vanillin production was selected.
- To check the feasibility of the process design proposed in existing literature using computational tools such as simulation software (AspenPlus V10) and short-cut tools like DSE (Distillation Sequence Efficiency).
- To implement a simulated process in AspenPlus both for the base case design selected form the bibliography and for the alternatives proposed in the present work.
- To study the effect of different thermodynamical physicochemical property estimation method selection in AspenPlus, i.e. NRTL and UNIQUAC.
- To conduct a solvent screening process for conventional solvents as well as for aliphatic alcohols C_6-C_8 as promising alternatives. This solvent screening process must take into account both energetic and environmental impact metrics.
- To optimize the value of the parameters and degrees of freedom available in the systems under study for the most promising process design.

3. METHODOLOGY

3.1. PROBLEM DEFINITION

First of all, a search is conducted in the database Reaxys to find a suitable bioprocess for study and simulation. Reaxys (Elsevier) is a web-based tool that retrieves chemical data from published literature, including journals and patents.

The information that can be retrieved includes chemical properties / reactions as well as experimental procedures and protocols from journals and scientific papers. It includes current and historical data from organic, inorganic and organometallic chemistry.

One selected prerequisite prior to conducting the search was that the selected process had to use a renewable source material as feed, which is important for the overall viability and profitability of a bioprocess.

Other criteria important to the search conducted:

- The existence of available simulation-based bibliography (preferably in AspenPlus).

- That viable alternatives could be conceived from reading up-to-date bibliography, so that the simulations performed could be of interest today.

Once an existing bioprocess is selected, the next step is to define the scope of the present study. The process is studied in the literature from feedstock to end products, and problem zones or *“hotspots*” for existing processes are singled out.

The energy requirement data of the unit operations involved in each section of the process is the primary criteria for the definition of such hotspots.

The dashed square (battery limit) in *Figure 9* shows the sections of the process selected for simulation .

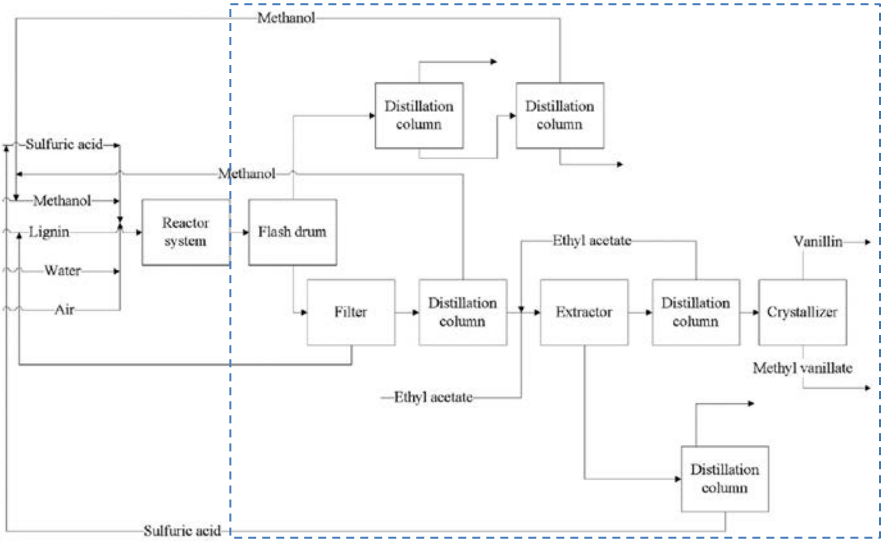


Figure 9. Scope of the present study simulations under a conventional vanillin production process *(based on Wongtanyawat et al., 2018).*

The raw materials used for vanillin production are the same for all proposed alternatives and include: Kraft lignin, methanol, water, sulfuric acid and air. The main product obtained is vanillin while methyl vanillate is a byproduct. Methanol and water are required in the acidic feed mixture when non-aliphatic alcohols are used as solvents. Sulfuric acid's function is to break down the lignocellulosic structure of Kraft lignin. Oxigen comes from the oxidation reaction in the form of air. Ethyl acetate and the solvents listed in *Table 1*, as well as aliphatic alcohols C_6-C_8 are used as solvents for the liquid-liquid extraction step, and a simulation is run for each and every one of them to asess their performance.

A design specification of 10 kg/h (99.5% commercial grade product purity) vanillin production is selected for implementation. This value ammounts for approximately 0,5% of global vanillin production worldwide from the petrochemical route. The reaction temperature and pressure of 210ºC and 48 bar are selected, respectively, based on existing data *(Wongtanyawat et al., 2018).* The pressure for the whole process afterwards is set at 1 bar.

3.2. BASE CASE DESIGN

The selected process is based on the most complete and reliable existing process data found in the available literature (*Wongtanyawat et al., 2018; Schorr et al.2014*).

It is decided that the process from *Figure 9,* excluding the reaction step, would be the base case design to be simulated. The simulation uses as feed the data from the reactor's outlet stream and has as its output the final product (vanillin) and subproduct (methyl vanillate) coming form the purification (crystalization) step. *Figure 10* shows the process flowsheet diagram implemented in AspenPlus.

The phase separation section in between includes a flash drum (FLASH) and two columns (DIST-2 & DIST-3) used to separate the gaseous components from liquid ones. It also includes a filter which separates the mixed products from the unconverted lignin (FILTER), a second column for the recovery and recirculation of methanol (DISTIL-1) and an extraction unit which separates the sulfuric acid from the products (LL-EXT). Then, a column for solvent recovery (DIST-4) prior to product purification in the crystallizer and another for sulfuric acid recovery (DIST-5) follows.

The simulation is performed first in simplified models like SEP and SEP2 in AspenPlus (infinite / infinite analysis) prior to rigorous simulation in the following step. The ∞/∞ analysis permits the study of the interrelation of the system streams prior to column design / specification. It can shed light on unfeasible regions, low limit values, multiplicity regions, discontinuities, control difficulties, recommendable operation conditions and column profile combinations without intensive and complicated calculations. The name of the model is derived form the assumption of an infinite reflux and number of stages.

This simplified model is used in the filter and distillation column implementation, which use SEP blocks with split fractions specified for each component. This way initial mass and energy balances are performed and we begin the understanding of the process without incurring in common convergence problems in the mathematics of more complex models like the MESH one (i.e. available in the Radfrac unit of AspenPlus). More specifically, DSTWU shortcut method is not used in this case because it is not suitable for simulating non-ideal mixtures, such as the present ones.

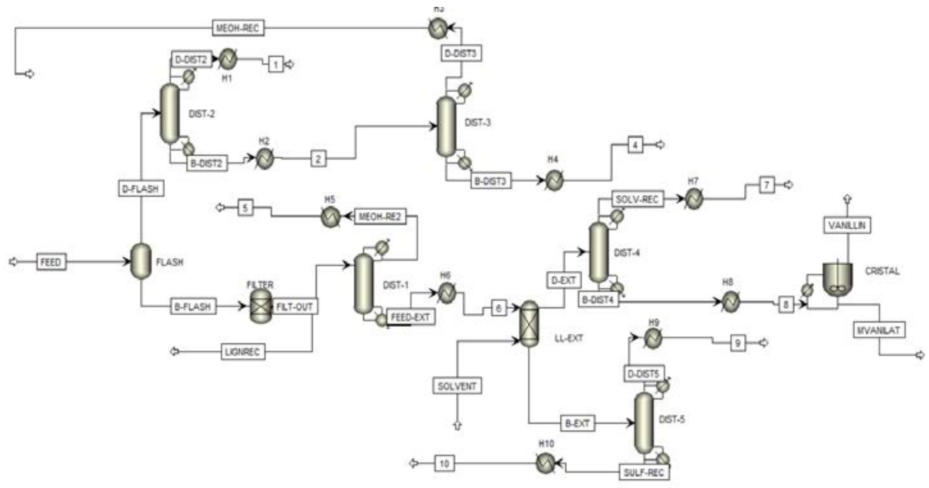


Figure 10. Flowsheet for base case design with simplified models in AspenPlus.

One exception to the use of ∞/∞ models for initial base case design assessment is the use of an Extract block for the liquid-liquid extraction unit operation. This is tested from the beginning in a rigorous block because of the crucial nature of the interactions between solvent (extractor) and the solute in the inlet stream.

The use of a simplified model with user specified split fraction separations could give initial false impressions about the applicability of certain extractive agents which wouldn't be corroborated once rigorous models are implemented.

Heater blocks are used for stream temperature estimations and a Crystallizer block is implemented for the purification process with working conditions of 5ºC of temperature and 1 bar of pressure, as per literature data *(Wongtanyawat et al., 2018*). A design specification of 99.5% vanillin is specified.

The results of the mass balance applied to the reactor assuming 122.02 kg/h of Kraft lignin feed provided to it provides the outlet stream of the reactor that corresponds to the feed of the separation train used in this study *(Table 2)* (*Wongtanyawat et al., 2018*).

Table 2. Feed mass balance (outlet from reactor) *(Wongtanyawat et al., 2018).*

Composition	Inlet (kg/h)	Outlet (kg/h)
Kraft lignin	*122.02*	*61.01*
Sulfuric acid	*61.01*	*35.39*
Air	*23,864.35*	*23,608.74*
Methanol	*6,053.07*	*5,750.42*
Water	*1,513.27*	*1,721.43*
Vanillin	-	*10.00*
Methyl vanillate	-	*2.03*
Dimethyl Ether	-	*139.11*
Carbon Dioxide	-	*266.24*
NO	-	*0.85*
SO_2	-	*18.52*

Most physical property data of the main components such as lignin and other chemicals are available in AspenPlus existing databanks. However, physical and chemical properties of methyl vanillate are not available, though, and are estimated introducing the compound's chemical formula in the molecule editor module of the software *(Figure 11)* and running the NRTL / UNIFAC component property estimation method, which calculates bonds and general physicochemical pure component properties derived from the molecular structure introduced, as well as binary interactions between components.

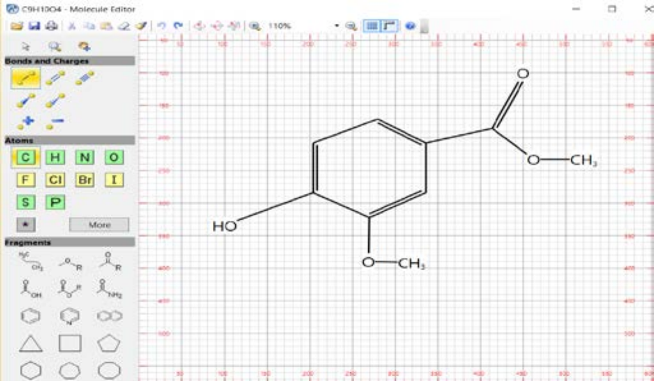


Figure 11. Methyl vanillate structure introduced in AspenPlus V10 molecule editor.

In general, NRTL + UNIFAC thermodynamical physical property method is selected for a base case design, with the exception of the crystallizer block which needs to use the ELECNRTL method for electrolyte property calculations. UNIQUAC method is selected for results comparison and to see the magnitude of the effect of choosing one method over another. For a more detailed description of physical method selection as a crucial step in process simulations see *Appendix 2*. The solvent used in the base case design is the most common one *(Wongtanyawat et al., 2018)*: ethyl acetate.

3.3. RIGOROUS PROCESS SIMULATION

Base case design includes the basic equipment design, which includes the specification of distillation columns such as the number of stages, feed location and reflux ratio. For this, following the ∞/∞ analysis model, a rigorous model consisting of Radfrac distillation columns is implemented. In this way, column designs are implemented and optimized. Energetic costs and CO_2 emissions of column operation are also estimated for each column, and used as parameters involved in subsequent solvent screening processes.

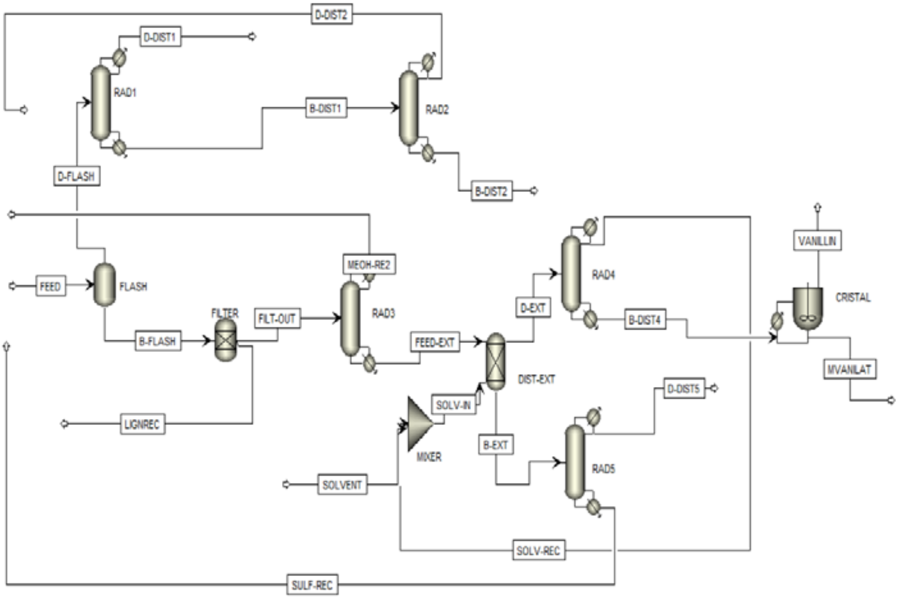


Figure 12. Flowsheet for the base case scenario rigorous simulation in AspenPlus.

Figure 12 shows the process flowsheet for the rigorous analysys of the base case design. The differences with the simplified model is that each distillation column consists of a Radfrac block and that heater blocks are not needed in this case, as temperature data is retrieved from the Radfrac models themselves. There exists a recirculation stream involving the recovered solvent (SOLV-REC).

3.3.1. Parameter optimization

The columns dispose of 5 degrees of freedom fulfilled by the following variables: reflux ratio, number of stages, feed stage, distillate flowrate and pressure.

To optimize the distillation columns, the reflux ratio influence is studied using a sensitivity analysis. Initially, a very high number of stages (e.g. 50) and a feed stage in the middle of the column (e.g. 25) are specified. The reflux ratio becomes correlated with the feed stage.

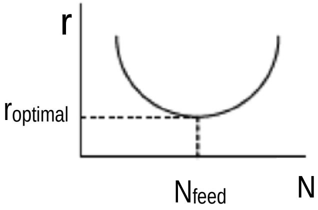


Figure 13. Sensibility analysis method for reflux ratio optimization example.

As the number of stages is very large then the minimum reflux obtained corresponds to the minimum reflux of the distillation column. According to heuristics, when multiplying the minimum reflux ratio by a factor of 1.2 and 1.5 we obtain a near optimal region for it.

By decreasing the number of stages and adjusting the feed stage where the reflux ratio is lowest, finally a reflux ratio fulfilling the range given by the heuristic is attained *(Figure 13)*. This procedure provides a near optimal number of stages and the optimal feed stage for the column.

3.4. ALTERNATIVE PROCESSES SIMULATION

Based on the literature (*Wongtanyawat et al., 2018*), there are principally two hotspots that are especially energetically demanding in the existing processes. Thus, they are the principal candidates for process improvement and alternative ideas brainstorming. Due to available data limitation in vanillin synthesis technology, process intensification on the reaction section is not considered.

The first hotspot is the liquid-liquid extraction unit plus its subsequent solvent recovery distillation column. This section has been intensively studied and simulated *(Kaygorodov et al., 2010; Wongtanyawat et al., 2018)* and different alternatives have been proposed, including the use of different solvents like benzene in a liquid liquid equilibrium (LLE) extraction scheme or the use of organic nanofilters or zeolites instead of liquid-liquid extraction for the separation of the sulfuric acid and water from the products. The use of benzene is problematic because of its toxicity and phase separation problems (*Table 1*).

The other prominent hotspot under study is the downstream separation columns that separate the gases from the liquid mixture and then recirculates the methanol to the system (RAD1 & RAD2 in *Figure 12*). In fact, previous simulations have found that this section is the most energy consuming one *(Wongtanyawat et al., 2018)*. Even knowing this, it is considered to be hard to find suitable replacements for this units as currently designed.

After initial simplified model analysis are performed, two alternative designs for this section are considered worthy of study.

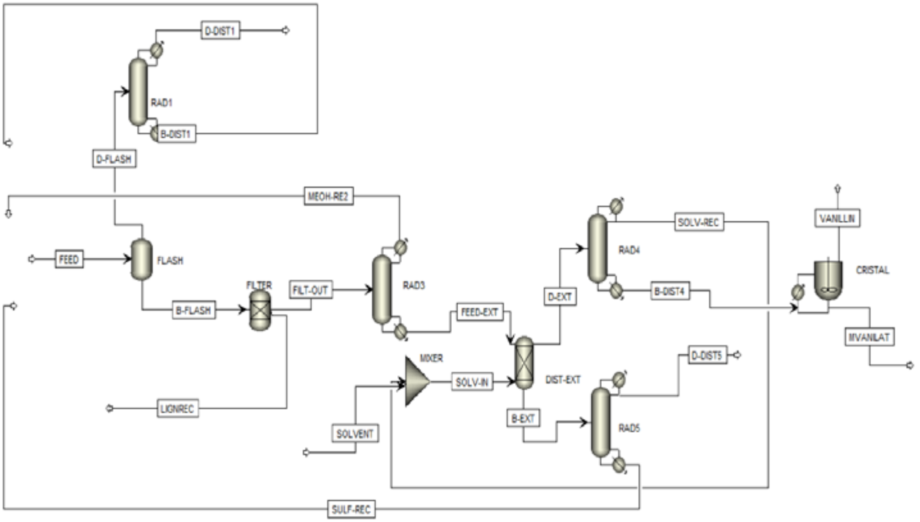


Figure 14. Flowsheet for Alternative 1 rigorous simulation.

Alternative one *(Figure 14)* involves the optimization of column RAD1 to have an acceptable output stream (B-DIST1) in terms of composition (especially the water / methanol ratio) as to make the second column present in the base case design (RAD2) not necessary.

This alternative is conceived thinking that water enters the system prior to the reaction step (*Figure 9*). Hence, in principle, a complete separation of water from the system in the RAD2 column might not be needed.

It seems that the only constraint is to maintain a constant ratio of methanol / water in the reactor feed. When this is fulfilled then the second distillation column becomes unnecessary, with a big overall reduction in energy consumption. The possibility of achieving this is studied using rigorous models.

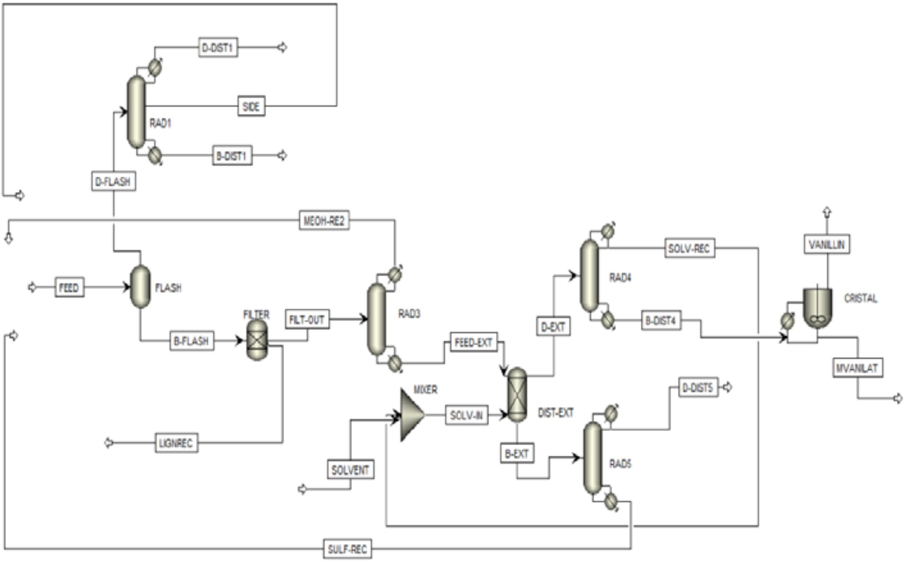


Figure 15. Flowsheet for Alternative 2 rigorous simulation.

Alternative two *(Figure 15)* involves the implementation of a side stream in the first phase separation column (RAD1) in order to reduce the amount of water recirculating to the reaction step (SIDE stream). This is a viable solution in case that alternative 1 is found to have an excess of water in the recycling stream that cannot be reduced via parameter fine-tuning. The ratio of methanol / water present in the recycling stream should in principle be more easily manipulated using this alternative and selecting the side stream stage, previous study of the composition profiles of the column provided by the software. This alternative also eliminates the need for a second column.

3.5. ALTERNATIVE PROCESS EVALUATION

Having proposed more than one alternative process design, a short-cut method is required to evaluate the performance of each alternative and relative to each other. Furthermore, the short-cut method avoids the use of rigorous simulations (at least in early stages of alternative evaluation) and their elevated computational and time requirements as well as convergence problems that normally appear and might need time to understand and solve adequately.

Bonet et al. (2015a) proposed as an optimization objective function for a process involving distillation columns the use of Carnot efficiencies when using the DSE (Distillate Sequence Efficiency) method. This method assumes that the energy consumed by the system is proportional to the streams flow rate and temperature differences between distillate and bottom streams *(eq. 1).*

$$DSE = \sum^{i} \frac{W_i}{F_c} \cdot \prod_{C} \eta_{ic} \tag{1}$$

With W_i being the flow rate of a stream, F_c being the crude feed flow rate and the multiplied term the Carnot efficiency of each column.

DSE is a method to determine the most energy efficient column sequence design possible. It is a method based on heuristics that assumes that a column is a heat engine that works between a hot source (reboiler of the column) and a cold source (condenser) that reduces entropy change instead of producing mechanical work. The method is calculated with only the information contained in the streams mass balances and temperature boiling points of a simulation ran under ∞/∞ simplified methods.

Distillate streams on W_i are multiplied by the Carnot efficiency of the column, while bottoms W_i streams are multiplied by one. When a recirculation stream is present, the multiplication term is composed of the efficiency of the columns inside the recycle minus one. Recirculations decrease the energy efficiency of the process (*Bonet et al., 2015).*

With this method, a pseudo-quantitative assessment of all possible alternatives is performed in a short time. Negative numbers may appear in the results obtained, which give us information about which system is effectively better in an energy efficiency kind of way but doesn't strictly correspond to an underlying physical magnitude (thus the pseudo-quantitative adjective applied to the method). It can be conceptualized as an energy-efficiency-based heuristic for fast and accurate selection of process design possibilities involving distillation columns.

When other unit operations that are not distillation columns are used in the process, correction factors are applied. A common way to define the efficiency of a liquid-liquid extraction unit, for example, could be to divide the mass flowrate of the solute product outlet stream in the unit divided by the mass inlet flowrate of said solute in the feed *(eq. 2).*

$$\mathrm{F}_{L-L} = \frac{w_{\mathrm{solute}_{\mathrm{out}}}}{w_{\mathrm{solute}_{\mathrm{in}}}} \qquad (2)$$

In the present work, the use of this correction factor is not needed, because the solute output stream (vanillin) is forced to have the 99.5% of the vanillin present in the inlet stream. Hence, when equation one is corrected with this design specification implemented, the correction factor amounts to a value around 1. The overall efficiency equation when correction factors are applied is the following (*eq. 3)*,

$$\mathrm{Efficiency}_{system} = DSE * \prod_{i}^{n} F \qquad (3)$$

With F implying a correction factor. It is clearly seen that with the present product purity design specification implemented that makes the correction factor for the liquid-liquid extraction unit close to 1, it has a negligible effect on the overall system efficiency. The efficiency of this unit is assessed by other means, as is explained in the solvent screening section.

The use of a proven method such as DSE is specially important considering that, in most processes, the most prominent factor in determining the cost of industrial implementation of a given process involving distillation columns is directly associated with the separation steps. The optimization of the column sequencing is, thus, an important step for economic viability.

3.5.1. DSE application example

Assume a flowsheet process design such as the one in *Figure 16.*

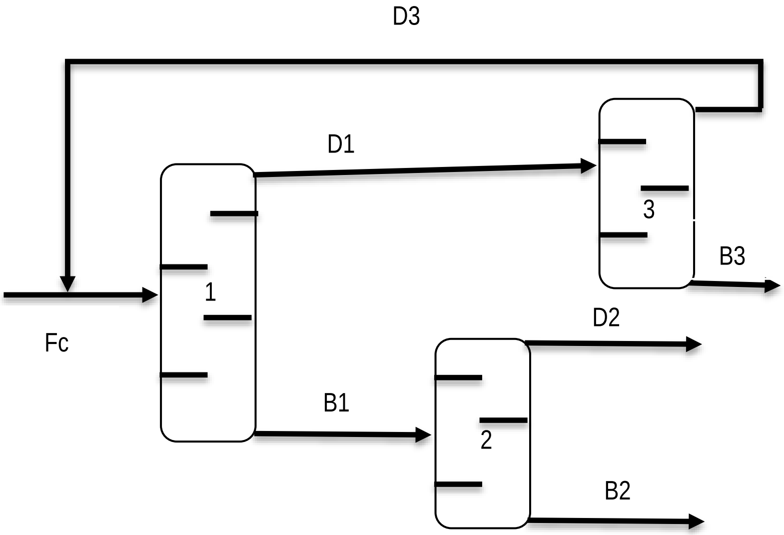


Figure 16. Method for DSE calculation using Carnot efficiencies example.

Its corresponding DSE, based on the type of streams (distillate, bottoms or recirculation stream) as explained earlier, ,is shown in *eq. 4.*

$$\mathrm{DSE}_{example} = \frac{B2}{\mathrm{Fc}} + \frac{D2}{\mathrm{Fc}} * \eta 2 \ + \frac{B3}{\mathrm{Fc}} * \eta 1 \ + \frac{D3}{\mathrm{Fc}} * (\eta 1 * \eta 3 \ - 1) \tag{4}$$

$$\eta 1 = \frac{T_{B1} - T_{D1}}{T_{B1}}$$

$$\eta 2 = \frac{T_{B2} - T_{D2}}{T_{B2}}$$

$$\eta 3 = \frac{T_{B3} - T_{D3}}{T_{B3}}$$

3.5.2. DSE application for Base Case design

Following the previous rules and descriptive notation for the streams, equation 5 is obtained.

(5)

$$\mathrm{DSE}_{base} = \frac{D-DIST_1}{FEED_1} * \eta_1 + \frac{B-DIST_2}{FEED_1} + \frac{D-DIST_2}{FEED_1} * (\eta_2 - 1) + \frac{MEOH-RE_2}{FEED_1} * (\eta_3 - 1)$$
$$+ \frac{D-DIST_5}{FEED_1} * (\eta_5) + \frac{B-DIST_4}{FEED_1} + \frac{SOLV-REC}{FEED_1} * (\eta_4 - 1)$$

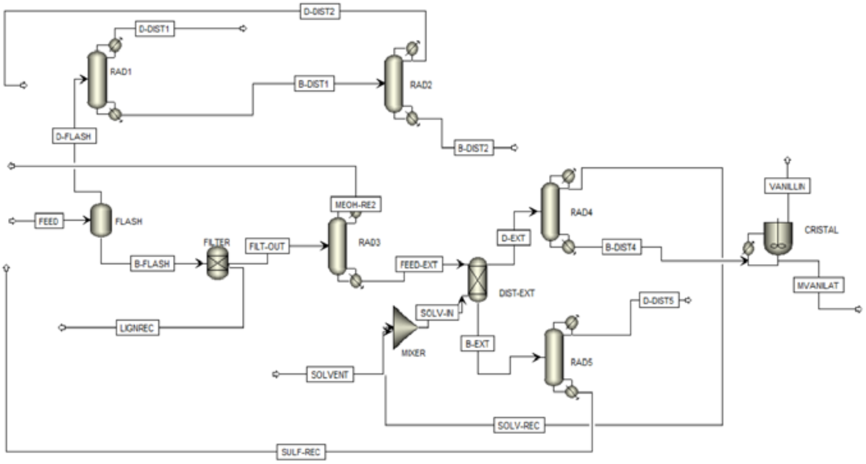


Figure 17. Flowsheet for base case (Radfrac model).

The recirculations that enter into the reactor (D-DIST2, MEOH-RE2 and SULFREC streams) in *Figure 17* are considered as recirculations into the system, even when the reactor is not modelled in this simulation.

LIGNREC stream recirculation is filtered solid unreacted lignin that isn't considered in this calculation.

Note: the efficiency of a bottoms stream is one, but also one is substracted when this bottoms stream is recirculated. Thus the term corresponding to a bottoms recycle stream becomes cancelled as happens here for the SULF-REC stream.

3.5.3. DSE application for Alternative Design 1

Based on the flowsheet of Fig. 16, then equation 6 is obtained.

(6)

$$\mathrm{DSE}_2 = \frac{D-DIST_1}{FEED_1} * \eta_1 + \frac{MEOH-RE_2}{FEED_1} * (\eta_3 - 1) + \frac{D-DIST_5}{FEED_1} * (\eta_5) + \frac{B-DIST_4}{FEED_1} + \frac{SOLV-REC}{FEED_1} * (\eta_4 - 1)$$

The B-DIST stream of the equation cancels itself in this case.

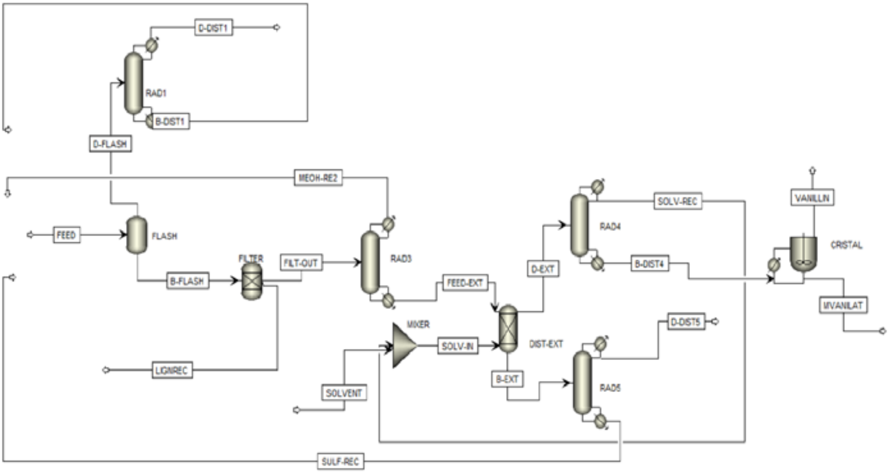


Figure 18. Flowsheet for Alternative 1 (Radfrac model).

3.5.4. DSE application for Alternative Design 2

For this case, based on the flowsheet of *Fig.17* , *equation 7* is obtained.

(7)

$$\mathrm{DSE}_3 = \frac{D-DIST_1}{FEED_1} * \boldsymbol{\eta_1}', + \frac{B-DIST_1}{FEED_1} + \frac{SIDE}{FEED_1} * (\eta_{SIDE} - 1) + \frac{MEOH-RE_2}{FEED_1} * (\eta_3 - 1) + \frac{D-DIST_5}{FEED_1} * (\eta_5)$$
$$+ \frac{B-DIST_4}{FEED_1} + \frac{SOLV-REC}{FEED_1} * (\eta_4 - 1)$$

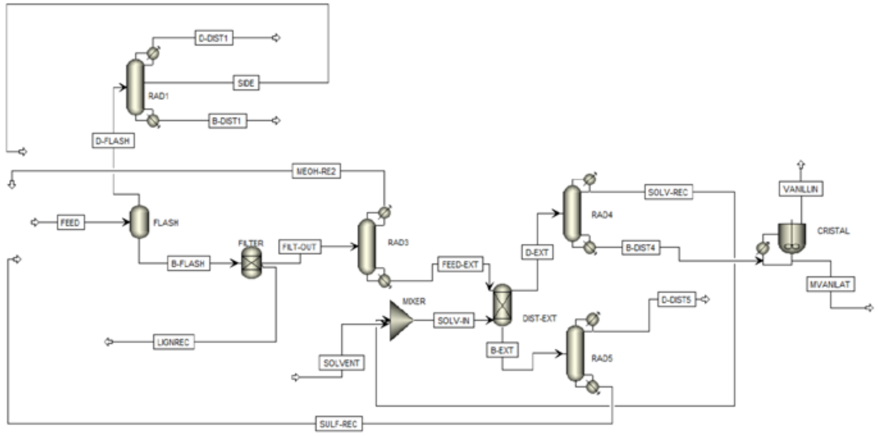
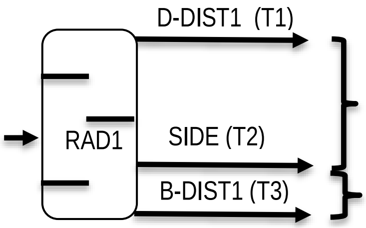


Figure 19. Flowsheet for Alternative 2 (Radfrac model).

With *Figure 19*'s Carnot efficiencies for column 1 being defined as *eq. 8, 9 and 10*.



$$\frac{T_2-T_1}{T_3} \qquad (8)$$

$$\eta_{SIDE} = \frac{T_3-T_2}{T_3} \qquad (9)$$

$$(8) + (9) = \eta_{1}, = \frac{T_3-T_1}{T_3} \qquad (10)$$

Figure 20. Carnot efficiencies for column 1 of Alternative Design 2 (Radfrac model). (All Carnot efficiencies must be calculated using Kelvin units for temperatures).

3.6. SOLVENT SCREENING

Apart from simulating those three alternatives mentioned, a solvent screening process is implemented to test the solvents present in *Table 1* as well as some others found in the literature as candidates for industrial application in the liquid-liquid extraction step of the process.

Liquid–liquid extraction is a unit operation that separates compounds (solutes) that are present in a stream when a suitable solvent that produces a phase split is added. A raffinate without the presence of solute is then collected, with the solute being placed in the extract stream. The solute is finally recovered mostly by distillation. Suitable solvent selection is the main difficulty to overcome in this unit operation.

For rapid evaluation of solvent alternatives in the early design stage of a process, two methods are selected based on *Bonet et al. 2014, 2015b.*

3.6.1. Minimum flow rate calculation

First, the minimum solvent flow rate needed for separation is an important parameter useful to compare the suitability of different alternative solvents. Once obtained it is used as an optimized input parameter for simulations.

The steps of the method are summarized as follows:

- First, a suitable thermodynamic method to estimate the compounds physicochemical properties must be selected. The model must be able to take into account the phase split. For most systems under study, UNIQUAC or NRTL are used, and/or UNIFAC when some data is not available in AspenPlus datasets.
- Second, a simple simulation involving a decanter unit must be performed *(Figure 21).* It doesn't model a physically real situation, but is a useful mathematical procedure. It consists of three input streams (solute, raffinate and solvent) and two output streams. Constant temperature and pressure are assumed (25ºC and 1 bar, respectively).

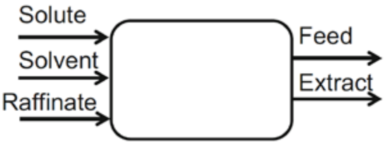


Figure 21. Decanter method for minimum solvent flowrate calculation *(Bonet et al. 2015b).*

- Third, the flow rate of the raffinate without solute is used as the calculation basis for the decanter input. The stream solute flowrate is calculated as the percentage of its specified recovery plus one and the result multiplied by its flow rate in the feed. The solvent flow rate is calculated using the design specification that satisfies the flow rate of solute in the feed which is an output stream of the decanter (using the Spec / Vary tool in AspenPlus).

3.6.2. Percentage of solvent lost in raffinate

The second method used for solvent screening involves the calculation of the percentage of lost solvent and it depends on its miscibility with the raffinate stream. A raffinate miscible solvent is also miscible in the feed.

The values are calculated according to the ratio of solvent flow rate in the feed stream in relation to the solvent stream flow rate.

When performing rigorous simulations, it is calculated as the quantity (in mass or moles) of the solvent present in the raffinate stream divided by the total quantity of solvent that entered the system. This data is obtained from the mass balances given by the simulations.

3.7. SOLVENTS POTENTIAL ENVIRONMENTAL IMPACT STUDY

In order to perform a complete assessment of possible solvents, not only energy usage metrics are evaluated but also each solvent potential environmental impact and waste production.

A useful parameter in this regard is the Aqueous Toxicity Parameter (ATP), which is defined as in *eq. 11*.

$$\text{ATP} = \frac{w_{makeup\ solvent}}{LC50} \tag{11}$$

LC50 [mg/L]: Fathead Minnow method

This method has the advantage that it takes into account the solvent intrinsic aqueous toxicity as a potential hazard. As a rule, a good solvent has both low solubility in the raffinate and a low Potential Environmental Impact (PEI) related with aqueous toxicity.

For these calculations, both ATP and PEI values, the waste reduction algorithm (*WAR, 2014)* is used.

For LC50 calculations, TEST software is the option of choice. Both are developed by the Environmental Protection Agency (EPA) of the USA.

3.7.1. The WAste Reduction algorithm (WAR)

It was developed so that the environmental impact of process designs could easily be evaluated. The goal of WAR is to reduce environmental and related human health impact at the design stage. The WAR algorithm evaluates processes in terms of potential environmental impacts (PEI). The potential environmental impacts of a chemical identified as a waste stream is defined as the effect that this chemical would have on the environment when it is simply emitted into it without atenuating measures.

The goal when using this tool is to minimize the PEI for a global process instead of minimizing the raw amount of waste and pollutants generated by a certain process.

The impact estimation algorithm has a complex model below the simple user interface surface, but is is flexible enough to allow users to emphasize / deemphasize different hazards as needed for particular applications, asigning different weights (if desired) to the perceived dominant hazard of a process.

WAR includes potential environmental impacts from eight categories:

- Human Toxicity Potential by Ingestion (HTPI)
- Human Toxicity Potential by Exposure (HTPE)
- Aquatic Toxicity Potential (ATP)
- Terrestrial Toxicity Potential (TTP)
- Global Warming Potential (GWP)
- Ozone Depletion Potential (ODP)
- Smog Formation Potential (SFP)
- Acidification Potential (AP)

In the present work, the ATP (Aquatic Toxicity Potential) category is the assumed most important one.

3.7.2. Toxicity Estimation Software Tool (TEST)

Developed to allow users to easily estimate the toxicity of chemicals using Quantitative Structure Activity Relationships (QSARs) methodologies. QSARs are mathematical models used to predict measures of toxicity from the physical characteristics of the structure of chemicals (known as molecular descriptors).

A chemical structure is introduced in the program, for example, by importing it from an included database of structures (i.e. a mol file). Once entered, the toxicity is estimated using one of the several available QSAR methods if experimental data is not available.

The required molecular descriptors are calculated within the software and no external programs or data are required. The results are compared against empirical data if available.

The model used in the present work for toxicity estimation is the 96-hour fathead minnow 50 percent lethal concentration (LC50).

4. RESULTS

4.1. ∞/∞ ANALYSIS

First the base case as specified in *Figure 10* is simulated using simplified models. SEP / SEP2 units with user-defined split fractions for each component are used to calculate mass balances and the separations taking place in the distillation columns of the process.

Heat X units are used to calculate stream temperatures. A Flash unit is also used, as well as an Extractor unit for the liquid-liquid extraction unit operation.

The viability of the separations as theoretically expected are proven, and the process is next updated to include rigorous simulations.

4.2. RIGOROUS PROCESS SIMULATION

The base case scenario is then migrated to rigorous mathematical models (MESH equations). In this case the Radfrac model is used for the distillation columns. The Extractor unit is conserved and a Crystallizer unit added for product purification *(Figure 12).*

4.2.1. Parameter Optimization

Table 3 values are obtained using the base case design as a template and the method explained in section *3.3.1.* and *Figure 13*.

Iterations are made until the reflux ratios obtained are inside the range of the minimum and maximum values obtained with the optimal reflux ratio heuristic.

Table 3. Parameter optimization for Base Case distillation columns (Radfrac model).

Parameter	Distillation column 1	Distillation column 2	Distillation column 3	Distillation column 4	Distillation column 5
Reflux Ratio	*1.37*	*2.10*	*1.64*	*1.30*	*1.35*
Number Stages	*4*	*12*	*10*	*10*	*8*
Feed Stage	*2*	*7*	*5*	*5*	*4*

4.2.2. Energy Consumption Analysis

Rigorous simulations give us energy consumption metrics such as the heat duty of the reboilers of the columns and of their top stage condensers.

The simulations are performed for each of the three possible designs under study. As can be seen in *Table 4,* the energy consumption of the first column of the base case scenario is the largest of them all, almost duplicating the magnitude of the other cases.

Column 2 only exists in the base case process design, so energy consumption is inexistent in the other ones. Column 4 also doubles the consumption of case 2 and is higher than case 3, while column 5 where sulfuric acid is recovered and recirculated presents a slightly lower consumption for the base case design.

Table 4. Heat duty for the reboilers of the distillation columns.

Case	Distillation column 1 [MW]	Distillation column 2 [MW]	Distillation column 3 [MW]	Distillation column 4 [MW]	Distillation column 5 [MW]	Total energy consumed [MW]
1. Base case	*10.96*	*5.45*	*0.14*	*0.04*	*0.27*	*16.87*
2. Alternative 1	*6.53*	*-*	*0.15*	*0.02*	*0.30*	*7.00*
3. Alternative 2	*6.85*	*-*	*0.14*	*0.03*	*0.29*	*7.31*

Table 5. Heat duty for the condensers of the distillation columns.

Case	Distillation column 1 [MW]	Distillation column 2 [MW]	Distillation column 3 [MW]	Distillation column 4 [MW]	Distillation column 5 [MW]	Total energy available [MW]
1. Base case	*-14.31*	*-5.39*	*-0.08*	*-0.04*	*-0.26*	*-20.08*
2. Alternative 1	*-9.87*	-	*-0.09*	*-0.01*	*-0.29*	*-10.27*
3. Alternative 2	*-10.26*	-	*-0.08*	*-0.03*	*-0.28*	*-10.65*

In *Table 5,* we see that for the base case scenario, columns 1 and 4 can give more energy than in cases 2 and 3, while column 3 and 5 can give a slightly lower amount.

4.3. ALTERNATIVE PROCESS EVALUATION

Based on the mass flow rates of the streams present in each process and the temperature estimations obtained in top / bottom stages of the columns, Carnot efficiencies are calculated for each column, in order to next obtain their DSE values *(Table 6).*

Table 6. Carnot efficiencies for the distillation columns of each process design.

Distillation column	Carnot Efficiency Descriptor	Base Case (5 columns)	Alternative 1 (4 columns)	Alternative 2 (4 cols. + side stream)
1	η_1	0.36	0.36	-
1’	$\eta_{1'}$	-	-	0.41
1’	η_{SIDE}	-	-	0.41
2	η_2	0.09	-	-
3	η_3	0.09	0.08	0.08
4	η_4	0.38	0.35	0.35
5	η_5	0.16	0.39	0.39

It is seen that efficiency for column 1 is slightly higher in alternative 2, for column 4 is slightly lower in the alternatives, and for column 5 is much higher in the alternative cases compared to the base case design.

To select the recirculation side stream stage in option 3 (alternative 2) that determines its temperature and consequently its Carnot efficiency, a composition profile plot is obtained *(Figures 22 & 23)*. As is shown in the results, if we want almost pure liquid methanol in our stream we need to specify the highest possible stage. The complication is that we need to recirculate pure methanol and some small degree of water that come from the top stages in the highest mass fractions possible but at the same time we need to purge the gases that also go out from the distillate stream (top stage), leaving the bottoms as a water purge stage. For this reason a partial condenser is used in the column, that in this way is defined as a tri-phasic (vapor-liquid-liquid) unit.

Lowering the side stream stage would diminish the side stream's Carnot efficiency because of the small temperature differences present between the side stream and the bottom stream when the side stream is specified as being in lower stages nearer the reboiler. This would also lower the DSE value, not to mention the harvest of methanol and the recovery of an excess amount of water.

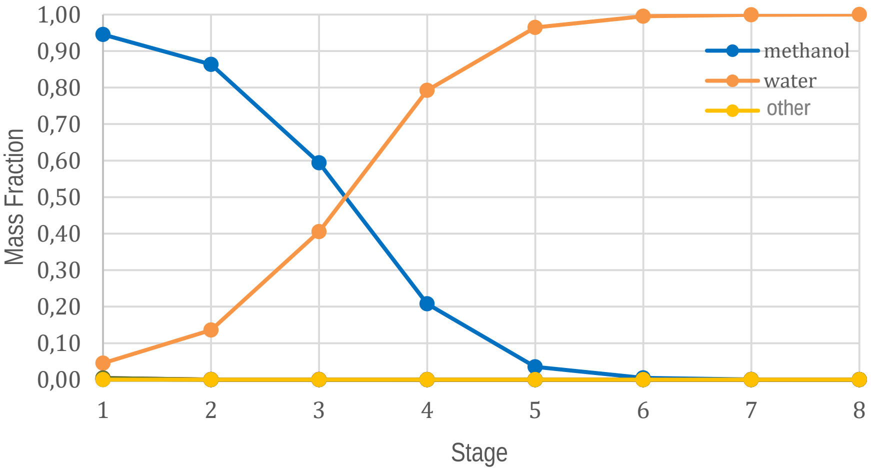


Figure 22. Composition profile for column 1 of Alternative design 2 (liquid phase).

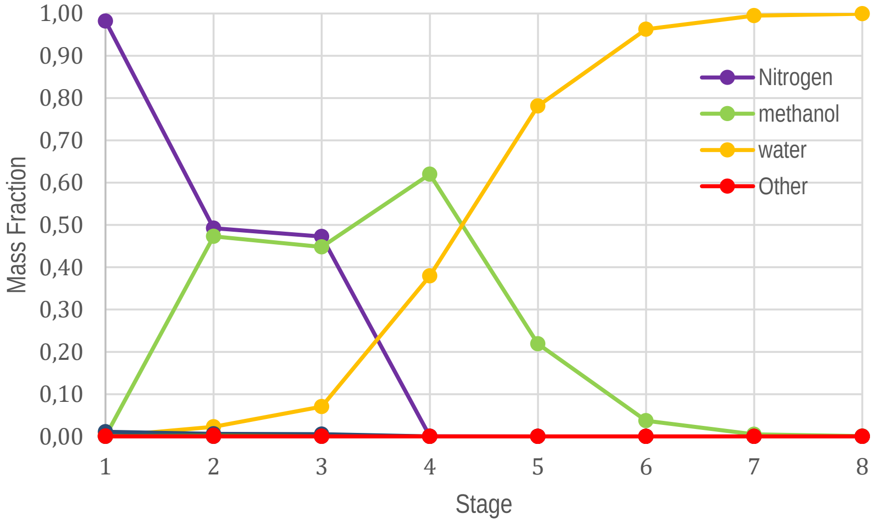


Figure 23. Composition profile for column 1 of Alternative design 2 (vapor phase).

Using these Carnot efficiencies, DSE values are calculated for each case *(Table 7).*

Table 7. DSE values and Relative Energy Comsumption for each process design proposed.

Case	Description	DSE [%]	Energy comsumption *relative to base case* ** based on Table 4*
1	Base case (5 columns)	15.57	-
2	Alternative 1 (4 columns)	26.99	0.41
3	Alternative 2 (4 cols. + side stream)	20.64	0.43

In general the process is not a highly efficient one. Nonetheless, the proposed alternatives do improve it. Alternative 1 is a 73% more energy efficient than the base case scenario based on the DSE values. Alternative 2 is a 32% more efficient.

As shown by these results, alternative 1 is the most energy efficient one, but another questions must be taken into consideration. Most important is to know if the composition of the recirculating stream coming from these different designs are all valid or the presence of water in one of them is too high as to render that process undesirable. To further our analysis of the three designs proposed, mass balances and fractions need to be studied.

The following analysis begins with the most complex design and ends with the simpler one (base case scenario). In *Table 8* we can see the mass balance of column 1' for alternative design 2.

Table 8. Mass balances for column 1' of Alternative Design 2 (4 cols. + side stream).

Component	D-Flash (inlet stream) [kg/h]	B-Dist1 (bottoms) [kg/h]	D-Dist1 (distillate) [kg/h]	Side (recirculating stream) [kg/h]
N_2	23608.40	0	23604.85	3.54
Methanol	5626.26	0.05	19.06	5607.16
Water	1529.19	647.12	0.10	881.97
DME	139.08	0	138.76	0.32
CO_2	266.22	0	265.96	0.26
NO	0.85	0	0.85	0
SO_2	18.50	0	16.76	1.74
Global	31188.50	647.18	24046.34	6494.99

It is shown that the bottoms stream is basically a water purge stream. In this process we do not want a complete water purge, though, but only to limit its amount in order to have a ratio methanol / water similar to the one entering the reactor. This is to be seen in the amount of water in the recirculating stream, which is far from null.

Also evident is that the gases are basically eliminated from the system in the distillate stream in gas phase via the partial condenser.

The side stream is basically composed of recirculated methanol. *Table 11* shows that the composition profile of the recirculating stream is 86% methanol, and 14% water. Here we have a H_2O / MeOH ratio of 16.3 while in the feed stream we have a value of 29.9. So , we have even less water and more methanol in the recirculating stream than in the feed, which might be desirable (the upper limit of water wanted being the feed stream ratio just mentioned). A 99.66% of methanol is recirculated this way, and a 57.67% of the water.

Table 9. Mass balances for column 1 of Alternative Design 1 (4 columns).

Component	D-Flash (inlet stream) [kg/h]	B-Dist1 (bottoms) [kg/h]	D-Dist1 (distillate) [kg/h]
N_2	23608.4	0	23608.4
Methanol	5626.26	5612.99	13.27
Water	1529.19	1529.17	0.013
DME	139.07	0	139.07
CO_2	266.22	0	266.22
NO	0.85	0	0.85
SO_2	18.50	0	18.50
Global	31188.50	7142.17	24046.3

Compared with alternative 2, which is the most complex design, alternative 1 has a relatively simpler composite profile *(Table 9).* Its recirculating stream (B-Dist1) recirculates almost all water and 99.76% of the methanol of the feed, a very slight improvement over alternative 2. It has the distinctive advantage that it eliminates completely the gases (because it has not a side stream near the distillate stage where the gases exit), not even containing the very small amount of certain ones that can be seen in alternative 2.

The diference with alternative 2 is that it recirculates almost 100% of the water, which may be an unwanted / wanted scenario depending on the acidification conditions we want to produce with this two-compound mix. Here we have a H_2O / MeOH ratio of 27,2 compared to the 29.9 of the feed stream. It is still a value below the limit imposed by that ratio and, in fact, it’s very close to it, which might be desirable if we want to recirculate a stream with the same ratio as the feed. But we must take into account that the feed stream might contain other acidic compounds that affect the overall acidity, so in this case we can’t say for sure without further studies if a more concentrated acidic recirculation (alternative 2) or a same ratio H_2O / MeOH but possibly with slightly less acidic conditions (the feed solution contains sulfuric acid) is the prefered solution (alternative 1).

Table 10. Mass balances for column 1 and 2 of Base Case design (5 columns).

Component	B-Dist1 (bottoms) [kg/h]	D-Dist1 (distillate) [kg/h]	B-Dist2 (bottoms) [kg/h]	D-Dist2 (recirculating stream) [kg/h]
N_2	0	23608.44	0	0
Methanol	5609.09	11.19	0.58	5608.51
Water	1532.43	0.08	1499.22	33.22
DME	0	139.07	0	0
CO_2	0	266.22	0	0
NO	0	0.85	0	0
SO_2	0.034	18.46	0	0.034
Global	7141.57	24044.33	1499.80	5641.76

Compared to the proposed alternatives, the base case design recirculates a 99.9% of the methanol with a H_2O / MeOH ratio of 0.006 compared to the 29.9 of the feed stream and very minor traces of SO_2 and no other gas present *(Table 10).* The obvious advantage is that it reduces the amount of recirculated water to almost none. But is it an advantage?

The premise of the proposed alternatives is that since water is present in the feed stream, the presence of water in the recirculating stream is not a problem to be dealt with. At least, if a good ratio of water to methanol can be achieved (as the previous data show) there doesn't seem to be necessary to purge all the water from the system.

Moreso, it doesn't seem optimal nor resource effective to have to constantly put water in the feed and have it completely removed in the first downstream separation stage.

Also, as we have seem, this complete water separation involves the use of an extra column. The Carnot efficiencies of columns 1 and 5 are also lower to the alternatives, as has been shown. Mass fraction data of the mass balances of the three designs under study are shown in *Table 11, 12 and 13*.

Table 11. Mass fractions for column 1 of Alternative Design 2 (4 cols. + side stream).

Component	D-Flash (inlet stream)	B-Dist1 (bottoms)	D-Dist1 (distillate)	Side (recirculating stream)
N_2	0.7570	0	0.9816	0.0005
Methanol	0.1804	0.0001	0.0008	0.8633
Water	0.0490	0.9999	0	0.1358
DME	0.0045	0	0.0058	0
CO_2	0.0085	0	0.0111	0
NO	0	0	0	0
SO_2	0.0006	0	0.0007	0.0003

Table 12. Mass fractions for column 1 of Alternative Design 1 (4 columns).

Component	D-Flash (inlet stream)	B-Dist1 (bottoms)	D-Dist1 (distillate)
N_2	0.7569	0	0.9818
Methanol	0.1804	0.7859	0.0005
Water	0.0490	0.2141	0
DME	0.0044	0	0.0058
CO_2	0.0085	0	0.0111
NO	0	0	0
SO_2	0.0006	0	0.0008

Table 13. Mass fractions for column 1 and 2 of Base Case design (5 columns).

Component	B-Dist1 (bottoms)	D-Dist1 (distillate)	B-Dist2 (bottoms)	D-Dist2 (recirculating stream)
N_2	0	0.98	0	0
Methanol	0.78	0.00046	0.00038	0.99
Water	0.21	0	0.99	0.006
DME	0	0.0058	0	0
CO_2	0	0.011	0	0
NO	0	0	0	0
SO_2	0	0.00077	0	0

4.4. SOLVENT SCREENING

4.4.1. Minimum flow rate calculation

A decanter simulation is performed for every solvent under scrutiny in AspenPlus V10 *(Figure 24).*

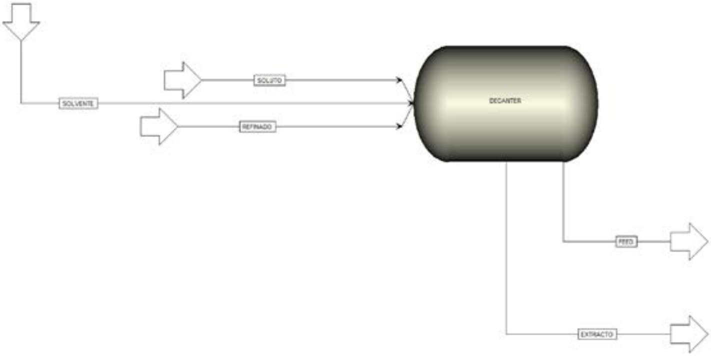


Figure 24. Flowsheet for decanter method simulation.

The minimum solvent flow rates where obtained for both UNIQUAC and NRTL physical property methods *(Table 14).*

Table 14. Minimum solvent flow rate values for each solvent

	Solvent	Minimum solvent flow rate [kg/h] (UNIQUAC)	Minimum solvent flow rate [kg/h] (NRTL)
1	Dichloromethane	37.53	-
2	Hexanol	38.17	47.22
3	Heptanol	47.08	103.70
4	Benzene	47.54	-
5	Octylamine	54.99	33.31
6	Octanol	57.51	87.20
7	Ethyl Acetate	60.65	-
8	Chloroform	61.02	4.39
9	Buthyl Acetate	61.58	68.55
10	Toluene	63.01	-
11	Hexane	68.14	-

NRTL method was selected as the starting one for its extended and versatile use. It was later discovered that it presents convergence problems when certain solvents are used.

Values are obtained with both methods in order to compare results when simulations with both methods converge. As can be seen in *Table 14*, even when no problems arise the values can be completely different between methods. The differences in results are solvent specific and no tendency can be appreciated.

The tendencies for UNIQUAC results are more consistent and reliable (see hexanol, heptanol and octanol values in *Table 14)*. It was concluded that NRTL method was not a valid one for the present study. Hence, UNIQUAC method was selected as a substitute for NRTL.

The importance of a thorough selection of a valid thermodynamical method in the first stages of a simulation was thus emphasized.

We see that the three solvents which need lower flow rates to achieve the specified separation of vanillin are dichloromethane, hexanol and heptanol. The two last ones are aliphatic alcohols, which have the advantages described in the introduction and could be used to simplify / compactify the process design further in future iterations.

The ones in 4th and 5th place are benzene and octylamine, respectively. They both have toxicity related problems, as will be seen in the next section.

The most commonly used extraction agent, ethyl acetate, is only in the 7th place of the 11 studied solvents.

4.4.2. Percentage of solvent lost in raffinate

It is calculated with two methods. Following the decanter method and through the mass balances obtained using the rigorous full-process simulations. This is done to cross-check the validity of the decanter method and to have a measure of the difference in values obtained.

As is shown in *Table 15,* the values obtained between methods are similar. No clear tendency is distinguished in the differences, as they are solvent–dependent and sometimes the values obtained with one method are higher and sometimes lower than the ones the other method gives. The magnitude of the difference between methods is also solvent-dependent. Octylamine presents convergence problems when used in the decanter method, so it's value is not included in *Table 15.*

Table 15. Amount of solvent lost in raffinate.

	Solvent	Solvent lost in raffinate [kg/h] (Rigorous simulation)	Solvent lost in raffinate [kg/h] (Decanter method)
1	Ethyl Acetate	17.34	20.89
2	Dichloromethane	4.13	4.78
3	Chloroform	2.47	2.95
4	Hexanol	1.60	2.01
5	Buthyl Acetate	1.68	1.91
6	Octylamine	1.36	-
7	Benzene	0.91	0.74
8	Heptanol	0.47	0.67
9	Toluene	0.32	0.27
10	Octanol	0.29	0.26
11	Hexane	0.02	0.01

Table 16. Percentage of solvent lost in raffinate.

	Solvent	Solvent lost in raffinate [%] (Rigorous simulation)	Solvent lost in raffinate [%] (Decanter method)	Relative Error
1	Ethyl Acetate	0.286	0.344	20.48
2	Dichloromethane	0.127	0.110	13.63
3	Hexanol	0.042	0.053	25.47
4	Chloroform	0.040	0.048	19.44
5	Buthyl Acetate	0.027	0.031	13.35
6	Octylamine	0.025	-	-
7	Benzene	0.019	0.016	18.63
8	Heptanol	0.010	0.014	42.26
9	Octanol	0.005	0.005	10.20
10	Toluene	0.005	0.004	17.48
11	Hexane	0.00024	0.00021	9.34

We can see that even when there are differences in the results that each method provides *(Table 16)*, the relative error between methods has a maximum magnitude of +/-25% in all cases except one (heptanol).

We can also see that the classifications in *Table 15 and 16 are* very similar, but *Table 16* takes into account not only the absolute values of the solvent lost in the raffinate but also the percentage lost relative to the minimum inflow of solvent needed to achieve the separation.

The obvious advantage of the decanter method is that it is based on a very fast and simple simulation that doesn't require a full rigorous simulation and is mostly free from convergence problems. It is, therefore, a good method for screening purposes.

Having calculated both solvent screening parameters, a first classification is proposed based on the overall solvent-specific resource requirements *(Table 17).*

Table 17. Resource-requirement solvent viability classification.

	Solvent viability classification	Minimum solvent flow rate [kg/h]	Solvent lost in raffinate [kg/h] (Rigorous simulation)
1	Hexanol	38.17	1.60
2	Dichloromethane	37.53	4.13
3	Heptanol	47.08	0.47
4	Benzene	47.54	0.91
5	Octylamine	54.99	1.36
6	Octanol	57.51	0.29
7	Buthyl Acetate	61.58	1.68
8	Toluene	63.01	0.32
9	Chloroform	61.02	2.47
10	Hexane	68.14	0.02
11	Ethyl Acetate	60.65	17.34

The classification on *Table 17 i*s done so that, for each solvent, its parameters sum value must be higher than the value of the preceding solvent and lower than the subsequent one.

Between the first two places, the ones that need a lower solvent flow rate, dichloromethane is much more soluble in the raffinate, so there is much more in the waste stream. This has implications for waste treatment / contamination assesment.

The third placed solvent, heptanol, needs a higher flow rate to achive a 99,5% separation of vanillin but it presents less than one third the solubility in the raffinate that hexanol presents.

Benzene and octylamine, as mentioned, present toxicity problems even when used industrially as is the case of benzene.

Ethyl acetate, as one of the most common solvents employed, is extremely soluble in the raffinate (and therefore is lost in the waste stream in great quantities) and needs a much higher flow rate than the first placed solvents, probably because of it.

Hexane has an extremely low solubility but, as we will see in the next section, an extremely high toxicity.

4.5. SOLVENT POTENTIAL ENVIRONMENTAL IMPACT STUDY

4.5.1. Aqueous Toxicity Parameter and LC50

In order to obtain a comprehensive classification of the viability of the studied solvents, environmental parameters regarding their toxicity need to be taken into account.

Using *eq. 11*, ATP values are calculated, and through the TEST software LC50 values obtained. We can see in *Table 18* that the LC50 and ATP values are inversely correlated.

The compound with the lowest toxicity is dichloromethane, followed by ethyl acetate and hexanol. The importance of using the ATP parameter in conjunction with a toxicity metric like LC50 is clearly seen when comparing the results for hexanol and ethyl acetate. If we only look at their LC50's we would think that ethyl acetate presents roughly half (49%) the toxicity of hexanol. Looking at their ATP values, though, we see that because hexanol needs a lower flow rate the "real" toxicity present in their respective waste streams is much closer (hexanol has a 26% more toxicity than ethyl acetate with the current values).

Table 18. ATP and LC50 of potential solvents.

	Solvent	Fathead minnow LC50 (96 h) [mg/L]	Aqueous Toxicity Parameter (ATP) [-]
1	Dichloromethane	319.93	0.12
2	Ethyl Acetate	230.18	0.26
3	Hexanol	117.07	0.33
4	Chloroform	96.14	0.63
5	Heptanol	36.17	1.30
6	Benzene	28.04	1.70
7	Toluene	34.24	1.84
8	Buthyl Acetate	17.99	3.42
9	Octanol	13.51	4.26
10	Octylamine	5.19	10.60
11	Hexane	2.5	27.26

We can also see that heptanol, even though it is much less soluble in the raffinate than hexanol, presents a much higher ATP (3.9 times higher).

A final classification is presented which takes into account both resource-related parameters, like solvent flow rate and presence of solvent in the waste stream (that equals the lost solvent in the raffinate), and the intrinsic toxicity of the extraction operation.

In *Table 19* we clearly see that significant changes are made in the classification of certain compounds compared to our previous classification. The algorithm for rating each solvent is done summing the minimum flow rate needed plus the solvent lost and multiplying the resulting quantity by their ATP value.

Especially striking is ethyl acetate’s case, which goes from the worst classified in the first classification to the third best when taking into account its low environmental impact compared to most of the other compounds. This is coherent with its use in industry.

Table 19. Solvent viability classification based on resource requirements and environmental hazards.

Actual classification (resources & environmental impact)	Previous classification (resource requirements)	Solvent	Minimum solvent flow rate [kg/h]	Solvent lost in raffinate [kg/h] (Rigorous simulation)	Aqueous Toxicity Parameter (ATP)
1	2	Dichloromethane	37.53	4.13	0.12
2	1	Hexanol	38.17	1.60	0.33
3	11	Ethyl Acetate	60.65	17.34	0.26
4	9	Chloroform	61.02	2.47	0.63
5	3	Heptanol	47.08	0.47	1.30
6	4	Benzene	47.54	0.91	1.70
7	8	Toluene	63.01	0.32	1.84
8	7	Buthyl Acetate	61.58	1.68	3.42
9	6	Octanol	57.51	0.29	4.26
10	5	Octylamine	54.99	1.36	10.60
11	10	Hexane	68.14	0.02	27.26

We also see that the aliphatic alcohols (hexanol, heptanol, octanol) present very different positions in our classification.

Hexanol is the second best when taking also into account toxicity metrics, while heptanol ranks fifth and octanol ninth. The chain length of these compounds, as we can see, greatly affects their viability.

4.5.2. Potential Environmental Impact (PEI) values

Using the WAR algorithm we obtained the total PEI indexes that use the 8 toxicity indicators explained in section *3.7.1* and the PEI ATP individual index that only takes into account the Aquatic Toxicity Potential of the mixture.

The difference between this PEI ATP value and the ATP value calculated previously is that in this case, the whole molar composition of the waste stream (D-DIST5) is included, so are their especific aquatic toxicities.

Each of the solvents used gives a different waste stream composition as some of them may be more or less miscible in the raffinate. Trace values of various components may be different among the streams too.

The PEI values obtained for the best five classified solvents in *Table 19* are shown in *Table 20*. The conclusions obtained can be different from the ones arrived at looking at *Table 19.* For example, hexanol has a lower PEI ATP than dichloromethane.

Table 20. Potential Environmental Index values for top solvents.

Classification in table 19	Solvent	Total PEI index [PEI/ h]	PEI ATP [PEI/ h]
1	Dichloromethane	0.576	0.003
2	Hexanol	0.324	0.001
3	Ethyl Acetate	1.255	0.006
4	Chloroform	0.318	0.002
5	Heptanol	0.094	0.001

4.6. ALIPHATIC ALCOHOLS VIABILITY AND PROPOSED NEW PROCESS DESIGN

The premise proposed in the introduction, that aliphatic alcohols C_6-C_8 could be good solutions as extractive vanillin agents in a liquid-liquid extraction operation is validated for hexanol and heptanol, while it isn’t the case for octanol. Nevertheless, we see that under the current classification, heptanol fares better than benzene (an industrially used solvent), and hexanol is the second best. Moreover, the minimum solvent flow rate requirements for hexanol are a lot lower than the requirements for the next best extractive agent (ethyl acetate), while it presents a close ATP value to it.

We conclude, then, that hexanol seems not only a viable but one of the best possible alternatives for vanillin extraction for the current process design.

Also, based on *Kaygorodov et al., 2010*, the acidification step needed for most solvents would be unnecessary for aliphatic alcohols. This opens the possibility that the selection of hexanol could help simplify the overall process in future designs. If we do not need to use methanol for creating a solution of certain acidity (*Wongtanyawat et al., 2018*), distillation column 3 could be removed, while distillation column one would only need to separate gases in its distillate stream and recirculate water through its bottom stream.

Overall we could get a more compact, intensified process that could be simpler and less energy consuming. The proposed process when hexanol is used is shown in *Figure 25*.

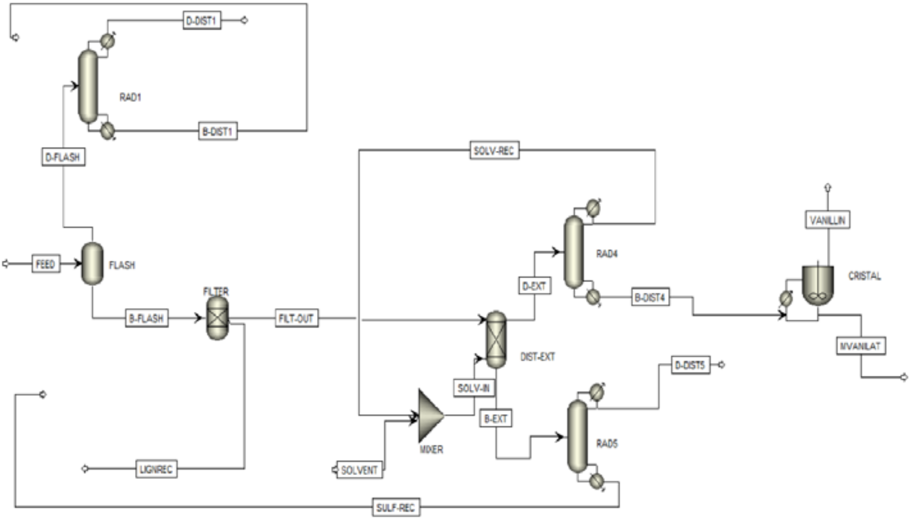


Figure 25. Flowsheet for simplified process design when using aliphatic alcohols as solvents .

The DSE value obtained for this design is higher than for the previous ones. More precisely, it's 38% higher than the second best option (alternative 1).

Lacking more information on the acidic conditions of the pre-reaction solution of the process, we can say that in principle the use of a design that uses hexanol as a liquid-liquid extraction agent is promising.

Because of its physicochemical properties, hexanol makes unnecessary the use of methanol as an extra acidification agent and permits a compactified design with only 3 distillation columns and an overall higher energetic efficiency *(Table 21).*

Table 21. DSE comparison chart including aliphatic-alcohols-exclusive process design option.

Case	Description	DSE [%]
1	Base case (5 columns)	15.6
2	Alternative 1 (4 columns)	27.0
3	Alternative 2 (4 cols. + side stream)	20.7
4	Hexanol (3 columns – No MeOH recovery step needed)	37.4

Rigorous simulations following this design gives us the energy consumption results shown in *Table 22 and 23*. It is shown that the reboilers of the aliphatic-alcohol-especific design consume much less energy that even alternative design 1. This is because the high mass flow rate of methanol that is no longer used (5750 kg/h; *Table 2*) has been taken out of the system, and the overall global mass circulation that needs phase separation is much lower.

Table 22. Heat duty for aliphatic-alcohols-exclusive process distillation columns.

Heat Duty	Distillation column 1 [MW]	Distillation column 2 [MW]	Distillation column 3 [MW]	Total energy Consumed / available [MW]
Reboiler	*3.38*	*0.060*	*0.33*	*3.77*
Condenser	*-5.34*	*-0.056*	*-0.32*	*-5.71*

Table 23. Energetic comparison chart

Base case (5 columns)	Energy consumed in reboilers [MW]	Difference between Consumed (reboilers) / available (condensers) energy [MW]
1. Base case (5 columns)	*16.87*	*-3.21*
2.Alternative 1 (4 columns)	*7.00*	*-3.26*
3.Alternative 2 (4 cols. + side stream)	*7.31*	*-3.34*
*4.*Alt. 3: aliphatic alcohols (3 cols.)	*3.77*	*-1.95*

Alternative 3's energy consumption is 22% that of the base case design, while *alternative 1* is 41% and *alternative 2* 43%.

5. CONCLUSIONS

- The use of simplified ∞/∞ models allowed us to assess the viability of the separations proposed in the literature without the need to perform rigorous computing-intensive simulations in the first stages of process analysis and design.
- Rigorous models and sensitivity analysis tools allowed us to optimize distillation column parameters such as the reflux ratio, number of stages and feed stage.
- The crucial importance of an appropiate physical properties estimation method in the first stages of a simulation is underscored. UNIQUAC proved to be a good method for the present system, while NRTL is not.
- DSE analysis allowed us a fast and reliable screening method for the proposed design alternatives in the first distillation columns of the downstream separation stage. The efficiency of the base case process is intrinsically low, while the proposed alternartives are higher. Alternative 1's DSE value is a 73% higher than the base case scenario. Alternative 2 is a 32% higher.
- Alternative 1 consumes 41% of the reboiler energy that consumes the whole base case process design, as proven by rigorous simulations. Alternative 2 consumes 43%.
- Solvent screening methods that take into account both the resource intensivity of each solvent and their raffinate solubility and related toxicity parameters are used in conjunction to achieve a comprehensive metric for solvent classification.
- The decanter method for solvent minimum flow rate proved to be a reliable and fast computationally non-demanding method.
- WAR algorithm and TEST software proved to be fast tools to retrieve toxicity related parameters when empirically available and to estimate them via QSAR methodology when not available.
- Solvent classification is different when all methods proposed are used as screening criteria as oposed to when only resource usage criteria is used. This

underscores the importance of also using toxicity metrics such as solvent solubility in the raffinate, ATP and LC50 parameters when classifying solvents to achieve a realistic evaluation of them.

- Aliphatic alcohols C_6-C_8 are viable solvents to achieve the liquid-liquid extraction step desired separations. Only hexanol and heptanol are considered good solvents, though.
- Hexanol achieved the best or second best results (depending on the weight given to selection criteria) in the solvent classification. The simulations tell us that it is less resource dependent and with a similar toxicity to the industrially used ethyl acetate and better than the also industrially used benzene.
- The viability of aliphatic alcohols as extraction solvents as proposed in the literature, which avoid the need of methanol usage in the pre-reactor feed preparation stage, may make the simplification / intensification of the process a possibility. A third design process alternative is thus proposed and it obtained the highest DSE efficiency among all considered options, being 239% higher than the base case value obtained. The global heat duty of this process design distillation columns is only 22% of the one of the base case scenario.
- In summary, the present methodology could be applied succesfully to a personally searched for and selected bioprocess of industrial interest which uses a renewable resource such as Kraft lignin as a feed stock. The results, as proven by rigorous simulations, have shed light on the possibility of energy comsumption reduction up to a 78% of the base case design, calculated as

$$\text{Energy improvement} = \frac{E_{base\ case} - E_{alternative}}{E_{base\ case}} * 100$$

Overall, 15 differents processes have been analyzed (11 solvents + 4 process designs, that were simulated using rigorous models).

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ACRONYMS

ATP	Aqueous Toxicity Parameter
DMC	Dichloromethane
DSE	Distillation Sequence Efficiency
EPA	Environmental Protection Agency
LC50	Lethal Dose 50
L-L	Liquid – Liquid
LLE	Liquid Liquid Equilibrium
MESH	Material-Equilibrium Summation-Heat
N	Number of stages of distillation column
N_{FEED}	Feed stage of distillation column
NRTL	Non-Random Two-Liquid
PEI	Potential Environmental Impact
QSAR	Quantitative Structure–Activity Relationship
RR	Reflux Ratio
TEST	Toxicity Estimation Software Tool
UNIFAC	Universal Functional group Activity Coefficient
UNIQUAC	Universal Quasichemical
WAR	Waste Reduction Algorithm
Wi	Mole flow rate of stream i
ηi	Carnot efficiency of column i

APPENDICES

APPENDIX 1: PRODUCTS OF ALKALINE OXIDATIVE LIGNIN DEPOLIMERIZATION

Table A1.1. Products of the Alkaline Oxidative Lignin Depolymerization Usually Studied in the Cases of Softwoods and Hardwoods *(Fache et al., 2016).*

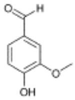
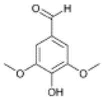
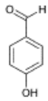
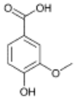
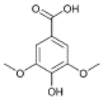
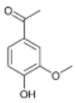
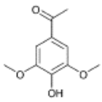
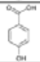
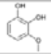
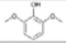
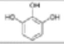
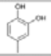
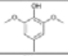
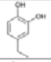
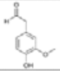
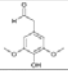
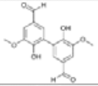
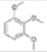
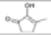
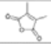
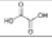
	Compound	Formula	Typical yields in softwoods	Typical yields in hardwoods
aldehydes	vanillin		6-12% 52, 53, 62-64	1-5% 50, 55, 62, 64-70
	syringaldehyde		0-0.7% 62	4-16% 50, 55, 62, 64-70
	p-hydroxy-benzaldehyde		0-0.5% 62, 64, 71	0-0.5% 50, 62, 64-66, 69
carboxylic acids	vanillic acid		0.5-1.5% 64	0.2-2.4% 55, 64, 67, 68, 70
	syringic acid		N.D.	0.5-3.9% 55, 64, 67, 68, 70
ketones	acetovanillone		0.6-6.4% 62, 71	0.3-2.6% 50, 62, 65, 66, 68
	acetosyringone		N.D.	1.5-4.2% 50, 62, 65, 66, 68

Table A1.2. Other Products of the Alkaline Oxidative Lignin Depolymerization Found in Literature *(Fache et al., 2016)*

	Compound	Formula
Phenolics	Phenol	
	Benzoic acid	
	Hydroxybenzoic acid	
	Guaiacol	
	Catechol	
	3-Methoxy catechol	
	Syringol	
	Pyrogallol	
	4-Methyl catechol	
	4-Methyl syringol	
	4-Ethyl catechol	
	2-(4-Hydroxy-3-methoxyphenyl)acetaldehyde	
	2-(4-Hydroxy-3,5-dimethoxyphenyl)acetaldehyde	
	Dehydrodivanillin	
	1,2-Bis(4-hydroxy-3-methoxyphenyl)ethane-1,2-dione	
Other molecules	1,2,3-Trimethoxy benzene	
	2-Hydroxy-3-methyl-2-cyclopentenone	
	3-Ethyl-2-hydroxy-2-cyclopentenone	
	3,4-Dimethyl maleic anhydride	
	Butyrolactone	
	Maleic acid	
	Fumaric acid	
	Succinic acid	
	Malonic acid	
	Propionic acid	
	Oxalic acid	
	Acetic acid	
	Formic acid	

APPENDIX 2: PHYSICAL PROPERTY METHOD SELECTION IN SIMULATORS

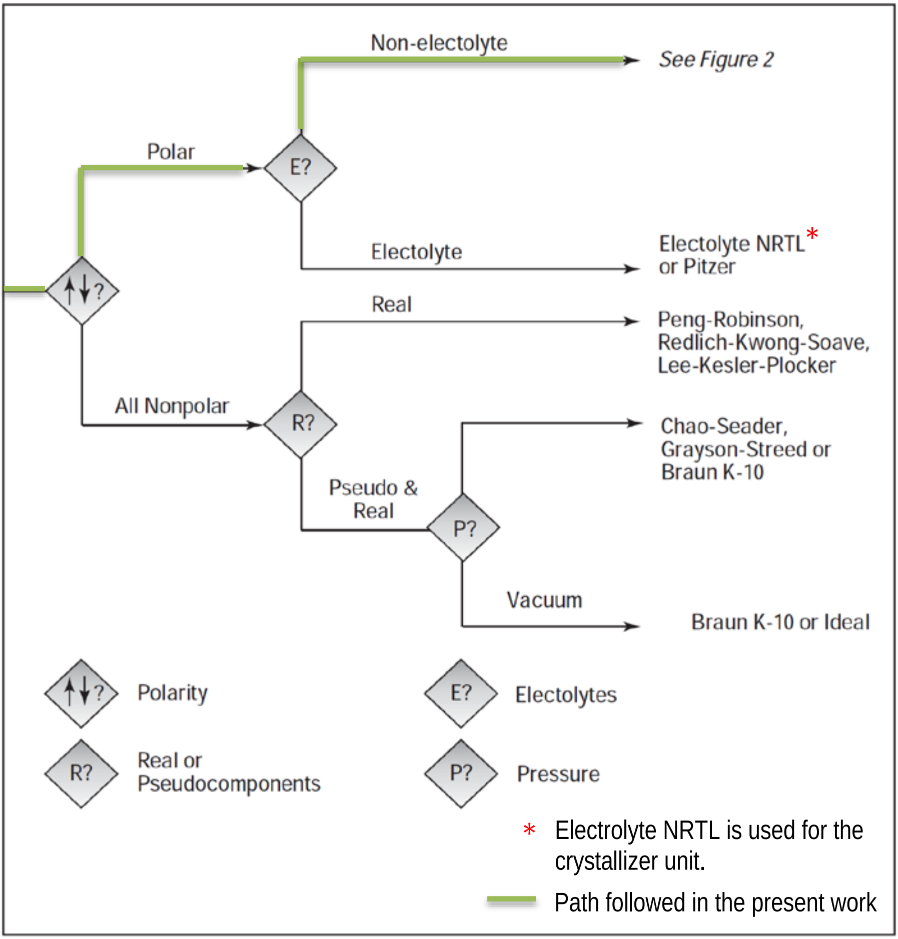


Figure A2.1. Decission tree for performing the the first steps for selecting physical property methods in simulation software. (*Carlson, C. Eric. Aspen Technology, Inc. "Don't Gamble With Physical Properties For Simulations". October 1996, Chemical Engineering Progress).*

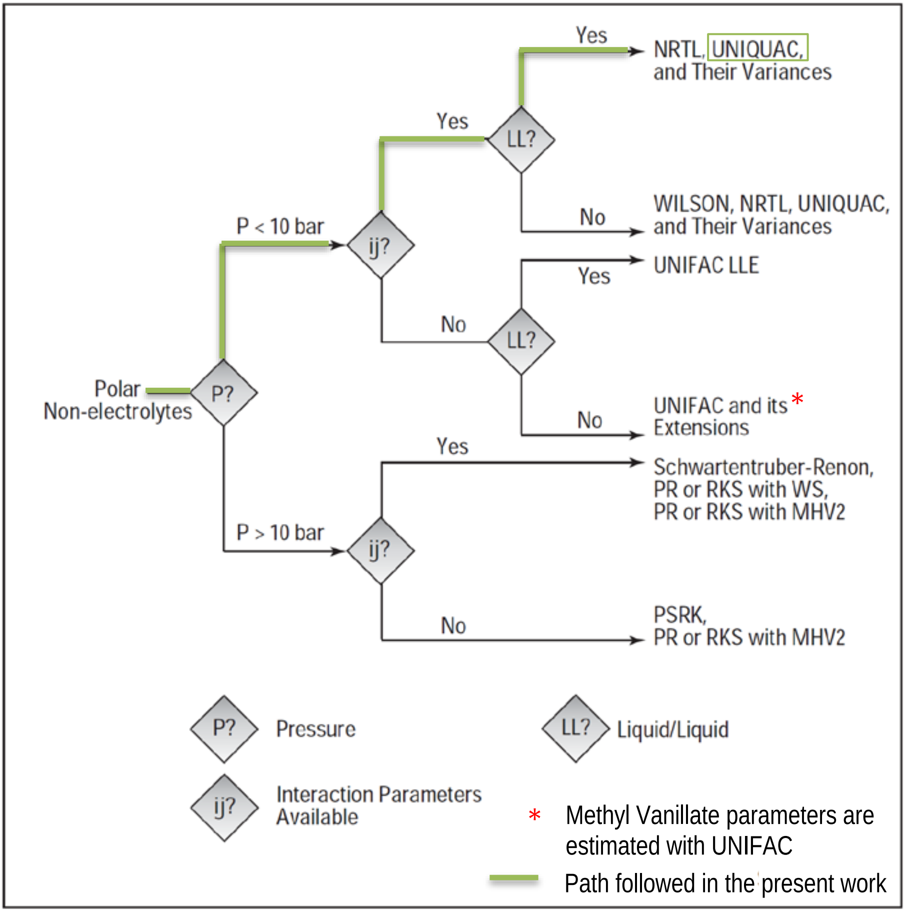


Figure A2.2. Decission tree for polar and nonelectrolyte components. *(Carlson, C. Eric. Aspen Technology, Inc. "Don't Gamble With Physical Properties For Simulations". October 1996, Chemical Engineering Progress).*

APPENDIX 3: COMPOSITION PROFILES

Table A5.1. Base case Rad1 distillation column liquid phase composition profile (mass basis).

Stage	H_2SO_4	N_2	MeOH	H_20	Vanillin	DME	CO_2	NO	SO_2	Methyl Vanillate
1	8.92 E-08	2.87 E-03	9.20 E-01	6.78 E-02	3.64 E-08	1.01 E-03	1.21 E-03	1.31 E-07	7.51 E-03	4.43 E-08
2	7.12 E-08	3.27 E-06	9.29 E-01	7.02 E-02	2.91 E-08	1.13 E-05	8.53 E-06	5.79 E-11	7.03 E-04	3.53 E-08
3	7.39 E-08	3.58 E-09	8.96 E-01	1.04 E-01	3.02 E-08	1.23 E-07	5.46 E-08	2.38 E-14	6.05 E-05	3.67 E-08
4	4.11 E-07	4.09 E-12	7.85 E-01	2.15 E-01	1.68 E-07	1.52 E-09	3.18 E-10	9.46 E-18	4.81 E-06	2.04 E-07

** each row / stage sums a total mass fraction of 1.*

Table A5.2. Base case Rad2 distillation column liquid phase composition profile (mass basis).

Stage	H_2SO_4	N_2	MeOH	H_20	Vanillin	DME	CO_2	SO_2	Methyl Vanillate
1	2.21 E-38	5.18 E-12	9.94 E-01	5.89 E-03	1.03 E-38	1.92 E-09	4.02 E-10	6.08 E-06	4.95 E-33
2	5.66 E-33	4.38 E-15	9.85 E-01	1.46 E-02	2.20 E-33	1.53 E-11	2.29 E-12	4.50 E-07	1.18 E-28
3	9.49 E-28	1.44 E-15	9.71 E-01	2.93 E-02	3.20 E-28	5.18 E-12	7.44 E-13	1.68 E-07	1.92 E-24
4	1.50 E-22	1.49 E-15	9.45 E-01	5.46 E-02	4.68 E-23	5.38 E-12	7.32 E-13	1.54 E-07	3.19 E-20
5	2.13 E-17	1.57 E-15	9.01 E-01	9.89 E-02	6.74 E-18	5.82 E-12	7.19 E-13	1.53 E-07	5.34 E-16
6	2.48 E-12	1.68 E-15	8.22 E-01	1.78 E-01	8.85 E-13	6.56 E-12	6.91 E-13	1.50 E-07	8.43 E-12
7	1.94 E-07	1.78 E-15	6.81 E-01	3.19 E-01	7.92 E-08	7.62 E-12	6.31 E-13	1.40 E-07	9.63 E-08
8	2.41 E-07	1.62 E-18	3.75 E-01	6.25 E-01	9.85 E-08	8.87 E-14	2.01 E-15	6.16 E-09	1.20 E-07
9	2.78 E-07	1.05 E-21	9.67 E-02	9.03 E-01	1.14 E-07	6.99 E-16	3.59 E-18	1.40 E-10	1.41 E-07
10	2.87 E-07	5.47 E-25	1.66 E-02	9.83 E-01	1.18 E-07	4.19 E-18	4.80 E-21	2.24 E-12	1.47 E-07
11	2.88 E-07	2.71 E-28	2.57 E-03	9.97 E-01	1.26 E-07	2.35 E-20	5.99 E-24	3.31 E-14	1.73 E-07
12	1.96 E-06	1.33 E-31	3.87 E-04	1.00 E+00	7.99 E-07	1.30 E-22	7.38 E-27	4.84 E-16	9.71 E-07

Table A5.3. Base case Rad3 distillation column liquid phase composition profile (mass basis).

Stage	H_2SO_4	N_2	MeOH	H_20	Vanillin	DME	CO_2	NO	SO_2	Methyl Vanillate
1	1.58 E-19	1.97 E-06	8.63 E-01	1.37 E-01	6.99 E-19	2.43 E-06	6.13 E-07	9.38 E-12	1.08 E-05	1.99 E-16
2	1.32 E-14	7.66 E-07	7.11 E-01	2.89 E-01	6.78 E-14	1.06 E-06	2.01 E-07	3.32 E-12	4.04 E-06	2.44 E-12
3	5.08 E-10	7.78 E-07	4.80 E-01	5.20 E-01	2.03 E-09	1.18 E-06	1.65 E-07	3.11 E-12	3.12 E-06	1.03 E-08
4	7.82 E-06	7.35 E-07	2.83 E-01	7.17 E-01	9.89 E-06	1.15 E-06	1.33 E-07	2.81 E-12	2.40 E-06	8.46 E-06
5	7.13 E-02	6.93 E-07	1.93 E-01	7.11 E-01	2.02 E-02	1.17 E-06	1.21 E-07	2.59 E-12	2.09 E-06	4.10 E-03
6	7.53 E-02	6.86 E-10	7.34 E-02	8.26 E-01	2.13 E-02	1.54 E-08	3.36 E-10	7.10 E-16	7.26 E-08	4.34 E-03
7	7.68 E-02	6.14 E-13	2.22 E-02	8.75 E-01	2.18 E-02	1.77 E-10	7.98 E-13	1.73 E-19	2.09 E-09	4.44 E-03
8	7.73 E-02	5.27 E-16	6.11 E-03	8.90 E-01	2.19 E-02	1.92 E-12	1.79 E-15	4.00 E-23	5.60 E-11	4.47 E-03
9	7.74 E-02	4.47 E-19	1.60 E-03	8.95 E-01	2.20 E-02	2.04 E-14	3.93 E-18	9.14 E-27	1.47 E-12	4.48 E-03
10	1.54 E-01	3.75 E-22	3.83 E-04	7.93 E-01	4.36 E-02	2.38 E-16	8.85 E-21	2.05 E-30	3.89 E-14	8.85 E-03

Table A5.4. Base case Rad4 distillation column liquid phase composition profile (mass basis).

Stage	H_2SO_4	MeOH	H_20	Vanillin	ETAC	Methyl Vanillate
1	1.43 E-15	5.38 E-04	5.14 E-01	1.08 E-09	4.85 E-01	4.85 E-01
2	1.06 E-11	1.36 E-04	9.94 E-01	2.09 E-07	5.44 E-03	5.44 E-03
3	2.19 E-08	6.36 E-05	9.98 E-01	1.12 E-05	2.16 E-03	2.16 E-03
4	4.54 E-05	5.77 E-05	9.97 E-01	5.94 E-04	2.15 E-03	2.15 E-03
5	1.29 E-01	6.66 E-05	7.17 E-01	1.23 E-01	5.87 E-03	5.87 E-03
6	1.29 E-01	1.42 E-05	7.22 E-01	1.23 E-01	1.32 E-04	1.32 E-04
7	1.29 E-01	3.01 E-06	7.22 E-01	1.23 E-01	2.95 E-06	2.95 E-06
8	1.29 E-01	6.38 E-07	7.22 E-01	1.24 E-01	6.60 E-08	6.60 E-08
9	1.43 E-01	1.38 E-07	6.67 E-01	1.57 E-01	1.82 E-09	1.82 E-09
10	1.54 E-01	3.83 E-04	7.93 E-01	4.36 E-02	4.79 E-11	8.85 E-03

Table A5.5. **Base case Rad5** distillation column liquid phase composition profile (mass basis).

Stage	*H_2SO_4*	*MeOH*	*H_20*	*Vanillin*	*ETAC*	*Methyl Vanillate*
1	4.50E-12	3.78E-04	9.00E-01	4.44E-11	9.97E-02	0
2	1.59E-08	5.48E-05	9.99E-01	3.96E-09	6.56E-04	0
3	2.94E-05	2.64E-05	1.00E+00	1.87E-07	2.68E-04	0
4	5.50E-02	2.39E-05	9.45E-01	1.14E-05	2.97E-04	0
5	5.50E-02	3.27E-06	9.45E-01	1.14E-05	2.09E-06	0
6	5.50E-02	4.46E-07	9.45E-01	1.14E-05	1.47E-08	0
7	6.81E-02	6.08E-08	9.32E-01	1.74E-05	1.06E-10	0
8	9.68E-01	1.13E-09	3.20E-02	1.98E-04	3.03E-12	0

Table A5.6. **Base case liquid-liquid extractor** composition profile (mole basis).

Stage	*H_2SO_4*	*MeOH*	*H_20*	*Vanillin*	*ETAC*	*Methyl Vanillate*
1	0.0384	0.0003	0.9264	0.0088	0.0253	0.0008
2	0.0363	0.0003	0.9338	0.0050	0.0245	0.0002
3	0.0351	0.0003	0.9379	0.0027	0.0241	0.0000
4	0.0345	0.0003	0.9399	0.0015	0.0238	0.0000
5	0.0342	0.0003	0.9411	0.0009	0.0236	0.0000
6	0.0340	0.0003	0.9418	0.0005	0.0234	0.0000
7	0.0339	0.0003	0.9422	0.0003	0.0233	0.0000
8	0.0339	0.0003	0.9424	0.0002	0.0232	0.0000
9	0.0338	0.0003	0.9426	0.0001	0.0232	0.0000
10	0.0338	0.0003	0.9427	0.0001	0.0232	0.0000

Table A5.7.**Alternative case 1 Rad1** distillation column liquid phase composition profile (mass).

Stage	*H_2SO_4*	*N_2*	*MeOH*	*H_20*	*Vanillin*	*DME*	*CO_2*	*NO*	*SO_2*	*Methyl Vanillate*
1	3.00E-19	2.26E-03	9.77E-01	9.52E-03	1.88E-17	1.06E-03	1.74E-03	1.17E-07	8.11E-03	1.91E-18
2	2.07E-13	4.20E-04	9.71E-01	2.79E-02	3.03E-12	4.11E-05	5.97E-05	7.03E-09	5.39E-04	3.96E-13
3	1.11E-07	3.91E-04	9.19E-01	8.05E-02	3.61E-07	4.14E-05	4.62E-05	6.36E-09	7.31E-05	6.00E-08
4	1.06E-07	4.35E-07	9.21E-01	7.89E-02	3.44E-07	6.77E-07	4.41E-07	2.86E-12	7.77E-06	5.73E-08
5	1.06E-07	4.76E-10	9.19E-01	8.11E-02	3.45E-07	1.10E-08	4.14E-09	1.26E-15	8.12E-07	5.74E-08
6	1.07E-07	5.23E-13	9.11E-01	8.88E-02	3.48E-07	1.85E-10	3.88E-11	5.59E-19	8.47E-08	5.79E-08
7	1.10E-07	5.79E-16	8.84E-01	1.16E-01	3.58E-07	3.48E-12	3.64E-13	2.48E-22	8.74E-09	5.96E-08
8	4.10E-07	6.55E-19	7.85E-01	2.15E-01	1.33E-06	8.95E-14	3.37E-15	1.11E-25	8.48E-10	2.22E-07

Table A5.8.**Alternative case 1 Rad3** distillation column liquid phase composition profile (mass).

Stage	*H_2SO_4*	*N_2*	*MeOH*	*H_20*	*Vanillin*	*DME*	*CO_2*	*NO*	*SO_2*	*Methyl Vanillate*
1	2.02E-19	1.47E-06	8.57E-01	1.43E-01	5.29E-16	1.01E-05	1.45E-06	1.02E-11	1.33E-05	6.24E-17
2	1.62E-14	5.60E-07	6.97E-01	3.03E-01	7.52E-12	5.81E-06	5.26E-07	3.79E-12	5.11E-06	1.09E-12
3	5.79E-10	5.72E-07	4.60E-01	5.40E-01	2.99E-08	7.68E-06	4.85E-07	3.79E-12	4.04E-06	5.06E-09
4	8.30E-06	5.55E-07	2.70E-01	7.30E-01	3.18E-05	7.85E-06	4.27E-07	3.60E-12	3.21E-06	5.94E-06
5	7.15E-02	5.21E-07	1.84E-01	7.20E-01	2.02E-02	7.40E-06	3.83E-07	3.31E-12	2.77E-06	4.11E-03
6	7.54E-02	4.70E-10	6.62E-02	8.33E-01	2.13E-02	2.39E-07	1.96E-09	1.06E-15	1.12E-07	4.33E-03
7	7.68E-02	3.86E-13	1.89E-02	8.78E-01	2.17E-02	6.70E-09	8.72E-12	3.05E-19	3.77E-09	4.41E-03
8	7.72E-02	3.06E-16	4.97E-03	8.92E-01	2.19E-02	1.78E-10	3.68E-14	8.41E-23	1.19E-10	4.44E-03
9	7.73E-02	2.40E-19	1.25E-03	8.95E-01	2.19E-02	4.66E-12	1.53E-16	2.29E-26	3.69E-12	4.45E-03
10	1.54E-01	1.81E-22	2.78E-04	7.93E-01	4.36E-02	1.20E-13	6.13E-19	5.93E-30	1.11E-13	8.85E-03

Table A5.9.**Alternative case 1 Rad4** distillation column liquid phase composition profile (mass).

Stage	H_2SO_4	*MeOH*	H_20	*Vanillin*	*ETAC*	*Methyl Vanillate*
1	6.48E-02	2.72E-04	1.35E-01	2.51E-02	7.72E-01	3.05E-03
2	7.70E-01	4.38E-06	3.12E-03	1.81E-01	2.34E-02	2.29E-02
3	7.65E-01	9.30E-07	6.30E-04	2.01E-01	5.60E-03	2.80E-02
4	7.40E-01	8.39E-07	5.60E-04	2.21E-01	4.95E-03	3.34E-02
5	7.16E-01	8.33E-07	5.53E-04	2.40E-01	4.87E-03	3.93E-02
6	6.81E-01	7.54E-09	6.55E-06	2.72E-01	9.12E-05	4.65E-02
7	6.32E-01	6.80E-11	7.68E-08	3.12E-01	1.68E-06	5.62E-02
8	5.74E-01	6.15E-13	8.91E-10	3.57E-01	3.04E-08	6.85E-02
9	5.09E-01	5.58E-15	1.03E-11	4.08E-01	5.41E-10	8.32E-02
10	4.38E-01	5.09E-17	1.18E-13	4.62E-01	9.46E-12	9.98E-02

Table A5.10. **Alternative case 1 Rad5** column liquid phase composition profile (mass).

Stage	H_2SO_4	*MeOH*	H_20	*Vanillin*	*ETAC*	*Methyl Vanillate*
1	4.05E-12	2.55E-04	9.03E-01	1.05E-10	9.67E-02	2.06E-11
2	1.42E-08	3.68E-05	9.99E-01	1.58E-08	6.34E-04	3.11E-09
3	2.63E-05	1.78E-05	1.00E+00	1.28E-06	2.60E-04	2.50E-07
4	4.92E-02	1.61E-05	9.50E-01	1.12E-04	2.84E-04	2.20E-05
5	4.92E-02	2.19E-06	9.51E-01	1.12E-04	1.97E-06	2.20E-05
6	4.92E-02	2.99E-07	9.51E-01	1.12E-04	1.36E-08	2.21E-05
7	5.41E-02	4.06E-08	9.46E-01	1.43E-04	9.47E-11	2.80E-05
8	9.41E-01	1.18E-09	5.64E-02	2.13E-03	3.56E-12	4.19E-04

Table A5.11. **Alternative case 1 liquid-liquid extractor** composition profile (mole basis).

Stage	*H_2SO_4*	*MeOH*	*H_20*	*Vanillin*	*ETAC*	*Methyl Vanillate*
1	0.0376	0.0002	0.9299	0.0064	0.0249	0.0011
2	0.0369	0.0002	0.9332	0.0045	0.0245	0.0008
3	0.0364	0.0002	0.9355	0.0033	0.0242	0.0005
4	0.0360	0.0002	0.9370	0.0024	0.0240	0.0004
5	0.0358	0.0002	0.9381	0.0018	0.0238	0.0003
6	0.0356	0.0002	0.9389	0.0013	0.0237	0.0002
7	0.0355	0.0002	0.9395	0.0010	0.0237	0.0002
8	0.0354	0.0002	0.9399	0.0008	0.0236	0.0001
9	0.0353	0.0002	0.9403	0.0006	0.0236	0.0001
10	0.0352	0.0002	0.9406	0.0004	0.0235	0.0001

Table A5.12.**Alternative case 2 Rad1** column liquid phase composition profile (mass).

Stage	*H_2SO_4*	*N_2*	*MeOH*	*H_20*	*Vanillin*	*DME*	*CO_2*	*NO*	*SO_2*	*Methyl Vanillate*
1	8.62 E-19	2.92 E-03	9.45 E-01	4.52 E-02	4.30 E-19	1.16 E-03	1.02 E-03	1.77 E-07	4.45 E-03	6.68 E-16
2	4.14 E-13	5.45 E-04	8.63 E-01	1.36 E-01	1.90 E-13	4.91 E-05	4.07 E-05	9.51 E-09	2.67 E-04	2.24 E-11
3	1.20 E-07	5.97 E-04	5.94 E-01	4.05 E-01	4.98 E-08	4.40 E-05	2.71 E-05	9.13 E-09	3.79 E-05	2.54 E-07
4	1.47 E-07	5.29 E-07	2.08 E-01	7.92 E-01	6.12 E-08	4.62 E-07	6.43 E-08	2.37 E-12	1.25 E-06	3.13 E-07
5	1.58 E-07	3.14 E-10	3.53 E-02	9.65 E-01	6.60 E-08	3.35 E-09	9.46 E-11	3.57 E-16	2.43 E-08	3.38 E-07
6	1.59 E-07	1.65 E-13	4.85 E-03	9.95 E-01	6.67 E-08	2.13 E-11	1.21 E-13	4.61 E-20	4.04 E-10	3.42 E-07
7	1.60 E-07	8.46 E-17	6.45 E-04	9.99 E-01	7.05 E-08	1.33 E-13	1.50 E-16	5.80 E-24	6.53 E-12	3.97 E-07
8	2.85 E-06	4.34 E-20	8.48 E-05	1.00 E+00	1.19 E-06	8.26 E-16	1.87 E-19	7.27 E-28	1.05 E-13	6.05 E-06

Table A5.13.Alternative case 2 Rad3 column liquid phase composition profile (mass).

Stage	*H_2SO_4*	*N_2*	*MeOH*	*H_20*	*Vanillin*	*DME*	*CO_2*	*NO*	*SO_2*	*Methyl Vanillate*
1	1.04E-19	2.75E-06	7.60E-01	2.40E-01	5.91E-18	2.88E-06	5.37E-07	1.91E-11	9.15E-06	4.32E-14
2	6.94E-15	1.06E-06	5.04E-01	4.96E-01	2.77E-13	1.16E-06	1.56E-07	6.42E-12	3.00E-06	9.93E-11
3	2.30E-10	9.76E-07	2.65E-01	7.35E-01	2.85E-09	1.16E-06	1.25E-07	5.07E-12	2.24E-06	5.93E-08
4	4.67E-06	9.07E-07	1.65E-01	8.35E-01	9.31E-06	1.11E-06	1.11E-07	4.38E-12	1.95E-06	1.65E-05
5	7.16E-02	8.71E-07	1.32E-01	7.72E-01	2.02E-02	1.06E-06	1.06E-07	4.47E-12	1.81E-06	4.11E-03
6	7.43E-02	9.38E-10	4.00E-02	8.60E-01	2.10E-02	1.35E-08	2.84E-10	1.35E-15	6.35E-08	4.27E-03
7	7.50E-02	9.32E-13	1.03E-02	8.89E-01	2.12E-02	1.57E-10	6.91E-13	3.69E-19	1.98E-09	4.31E-03
8	7.52E-02	9.03E-16	2.49E-03	8.97E-01	2.13E-02	1.77E-12	1.63E-15	9.74E-23	5.97E-11	4.32E-03
9	7.53E-02	8.70E-19	5.85E-04	8.98E-01	2.13E-02	1.99E-14	3.81E-18	2.55E-26	1.78E-12	4.34E-03
10	1.55E-01	8.52E-22	1.24E-04	7.92E-01	4.39E-02	2.16E-16	9.09E-21	7.64E-30	5.24E-14	8.90E-03

Table A5.14.Alternative case 2 Rad4 column liquid phase composition profile (mass).

Stage	*H_2SO_4*	*MeOH*	*H_20*	*Vanillin*	*ETAC*	*Methyl Vanillate*
1	6.56E-22	3.69E-05	7.93E-02	5.48E-19	9.21E-01	3.04E-19
2	9.11E-17	1.22E-05	4.63E-02	1.80E-14	9.54E-01	8.01E-15
3	6.24E-12	6.47E-06	3.08E-02	2.99E-10	9.69E-01	1.05E-10
4	4.01E-07	5.06E-06	2.44E-02	4.69E-06	9.76E-01	1.29E-06
5	2.08E-02	4.69E-06	2.22E-02	5.86E-02	8.86E-01	1.29E-02
6	1.98E-02	1.26E-06	1.08E-02	5.57E-02	9.01E-01	1.23E-02
7	1.93E-02	3.17E-07	5.01E-03	5.43E-02	9.09E-01	1.20E-02
8	2.10E-02	7.77E-08	2.26E-03	5.93E-02	9.04E-01	1.31E-02
9	1.60E-01	1.64E-08	6.86E-04	4.23E-01	3.21E-01	9.44E-02
10	2.24E-01	2.48E-10	1.04E-05	6.30E-01	7.85E-03	1.39E-01

Table A5.15. **Alternative case 2 Rad5** column liquid phase composition profile (mass).

Stage	*H_2SO_4*	*MeOH*	*H_20*	*Vanillin*	*ETAC*	*Methyl Vanillate*
1	2.13E-03	1.35E-04	8.99E-01	8.93E-05	9.89E-02	3.35E-08
2	8.85E-01	3.94E-06	1.03E-01	8.71E-03	3.35E-03	2.64E-06
3	9.87E-01	1.77E-07	3.60E-03	8.94E-03	3.27E-04	2.59E-06
4	9.88E-01	1.34E-07	2.18E-03	9.37E-03	2.51E-04	2.64E-06
5	9.88E-01	9.75E-10	3.73E-05	1.17E-02	4.69E-06	3.27E-06
6	9.85E-01	7.12E-12	6.41E-07	1.49E-02	8.82E-08	4.09E-06
7	9.81E-01	5.22E-14	1.10E-08	1.90E-02	1.66E-09	5.16E-06
8	9.76E-01	3.84E-16	1.88E-10	2.43E-02	3.12E-11	6.54E-06

Table A5.16. **Alternative case 2 liquid-liquid extractor** composition profile (mole basis).

Stage	*H_2SO_4*	*MeOH*	*H_20*	*Vanillin*	*ETAC*	*Methyl Vanillate*
1	0.0348	0.0001	0.9316	0.0081	0.0245	0.0008
2	0.0348	0.0001	0.9332	0.0073	0.0242	0.0004
3	0.0348	0.0001	0.9344	0.0066	0.0239	0.0002
4	0.0347	0.0001	0.9354	0.0060	0.0237	0.0001
5	0.0347	0.0001	0.9363	0.0054	0.0235	0.0001
6	0.0347	0.0001	0.9370	0.0049	0.0233	0.0000
7	0.0346	0.0001	0.9378	0.0044	0.0231	0.0000
8	0.0346	0.0001	0.9385	0.0039	0.0229	0.0000
9	0.0346	0.0001	0.9391	0.0034	0.0228	0.0000
10	0.0346	0.0001	0.9398	0.0029	0.0226	0.0000
11	0.0346	0.0001	0.9404	0.0025	0.0224	0.0000
12	0.0346	0.0001	0.9411	0.0020	0.0223	0.0000
13	0.0345	0.0001	0.9418	0.0015	0.0221	0.0000
14	0.0342	0.0001	0.9427	0.0010	0.0219	0.0000
15	0.0316	0.0001	0.9465	0.0005	0.0213	0.0000

Table A5.17.**Alternative case 3 (aliphatic alcohols)** Rad1 column liquid phase composition profile (mass).

Stage	*H_2SO_4*	*N_2*	*H_20*	*Vanillin*	*DME*	*CO_2*	*NO*	*SO_2*	*Methyl Vanillate*
1	7.10E-14	5.47E-03	3.55E-02	5.26E-12	5.14E-01	1.32E-02	7.06E-07	4.32E-01	4.96E-12
2	6.66E-07	2.73E-04	9.72E-01	2.66E-06	5.85E-03	3.80E-05	4.55E-09	2.18E-02	1.07E-06
3	6.37E-07	3.42E-08	1.00E+00	2.57E-06	1.37E-05	1.68E-08	1.93E-13	1.67E-04	1.06E-06
4	2.76E-06	7.84E-12	1.00E+00	1.10E-05	4.72E-08	1.18E-11	1.37E-17	1.72E-06	4.43E-06

** the number of stages for Rad 1 column could be reduced in this process design*

Table A5.18.**Alternative case 3 (aliphatic alcohols)** Rad4 column liquid phase composition profile (mass).

Stage	*H_2SO_4*	*H_20*	*Vanillin*	*ETAC*	*Methyl Vanillate*
1	4.89E-03	1.73E-01	2.78E-03	8.17E-01	3.24E-04
2	5.10E-01	7.26E-03	9.08E-02	3.83E-01	9.43E-03
3	7.98E-01	8.70E-04	1.35E-01	5.05E-02	1.54E-02
4	8.06E-01	6.38E-04	1.51E-01	2.25E-02	1.93E-02
5	7.89E-01	6.20E-04	1.66E-01	2.02E-02	2.37E-02
6	7.74E-01	7.58E-06	1.95E-01	1.54E-03	2.93E-02
7	7.33E-01	9.13E-08	2.30E-01	1.11E-04	3.70E-02
8	6.80E-01	1.09E-09	2.72E-01	7.64E-06	4.74E-02
9	6.18E-01	1.29E-11	3.21E-01	4.91E-07	6.07E-02
10	5.47E-01	1.50E-13	3.76E-01	2.90E-08	7.69E-02

Table A5.19. Alternative case 3 (aliphatic alcohols) Rad5 column liquid phase composition profile (mass).

Stage	H_2SO_4	H_20	*Vanillin*	*ETAC*	*Methyl Vanillate*
1	3.64E-12	9.91E-01	1.04E-10	8.00E-03	7.58E-13
2	1.13E-08	1.00E+00	8.18E-09	1.71E-04	2.11E-11
3	2.07E-05	1.00E+00	3.79E-07	7.23E-05	3.53E-10
4	3.78E-02	9.62E-01	2.08E-05	8.07E-05	7.38E-09
5	3.78E-02	9.62E-01	2.08E-05	2.03E-06	7.38E-09
6	3.78E-02	9.62E-01	2.09E-05	5.11E-08	7.39E-09
7	3.94E-02	9.61E-01	2.23E-05	1.29E-09	7.77E-09
8	8.80E-01	1.20E-01	4.80E-04	6.86E-10	1.67E-07

Table A5.20. Alternative case 3 (aliphatic alcohols) liquid-liquid extractor composition profile (mole basis).

Stage	H_2SO_4	H_20	*Vanillin*	*ETAC*	*Methyl Vanillate*
1	0.02822	0.96485	0.00339	0.00239	0.00037
2	0.02776	0.96701	0.00222	0.00209	0.00015
3	0.02751	0.96825	0.00150	0.00193	0.00007
4	0.02736	0.96900	0.00103	0.00183	0.00003
5	0.02726	0.96949	0.00072	0.00177	0.00001
6	0.02720	0.96982	0.00050	0.00173	0.00001
7	0.02716	0.97005	0.00035	0.00170	0.00000
8	0.02712	0.97021	0.00024	0.00168	0.00000
9	0.02709	0.97034	0.00017	0.00167	0.00000
10	0.02705	0.97045	0.00012	0.00166	0.00000
11	0.02696	0.97060	0.00008	0.00165	0.00000
12	0.02671	0.9411	0.0020	0.9363	0.0000