



Scaled-up and economic assessment approach of the split-phase glycolysis process for the recycling of flexible polyurethane foam wastes

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Abstract

The economic viability of the split-phase glycolysis process for the recycling of any kind of flexible polyurethane foam waste employing crude glycerol as cleavage agent has been demonstrated. First, experiments at pilot plant scale were carried out to check that the process can be extrapolated to larger scales. With the goal of scaling-up the process from laboratory scale to pilot plant, geometric similarity criteria were applied together with dynamic similarity for laminar flow in agitated tank reactors. Hence, a pilot plant installation was designed with geometrically similar equipment to those used for lab scale, obtaining analogous results in terms of recovered polyol properties. Then, the basic design of a split-phase glycolysis industrial plant with a capacity for treating 270 Tm per year of flexible PU foams scraps was proposed. Finally, the economic feasibility of such recycling process was confirmed because of the obtention of a Net Present Value (NPV) of 1,464,555€, with an Internal Rate of Return (IRR) of 27.99%, and a payback time between 4 and 5 years.

Keywords Glycolysis · Scale-up · Polyurethane wastes · Economic assessment · Circular economy · Crude glycerol

Introduction

Polyurethanes (PUs) are plastic polymers made by combining isocyanates and polyols. The global polyurethane market was valued at USD 76,600 million in 2020 and will reach USD 94,160 million by the end of 2027, growing at a compound annual growth rate (CAGR) of 3.5% during 2022–2027 [1]. The wide diversity of polyols and isocyanates allows the synthesis of numerous different compounds covering a huge range of applications. The polyurethanes are classified mainly in foams (flexible and rigid) and in the denominated CASEs (Coatings, Adhesives, Sealants, and Elastomers) [2]. On the other hand, the flexible foams can be separated into three main groups: conventional, viscoelastic and high resilience foams [3].

Nowadays, PU-recycling processes are one of the most important goals of the research and industrial worlds as a

direct result of the enforcement of the environmental legislations. The most important chemical recycling processes of polyurethanes have been widely commented by means of an exhaustive revision of the literature, highlighting the processes of glycolysis, aminolysis, hydrolysis, and phosphorolysis. However, most of these processes have not achieved the industrial and commercial scale, even in some cases the development grade is just laboratory tests [4].

There are some remarkable aspects relative to the development grade of the different available technologies of chemical recycling for polyurethanes. Glycolysis is the most important chemical recycling method for polyurethanes to date, with high expectation of being developed soon at industrial scale [4]. It is based on a transesterification reaction in which a hydroxyl group from a glycol replaces the ester group containing a carbonyl carbon of a urethane bond. This reaction produces polyols whose properties can be controlled to some extent and may be similar to those from the original material [5]. Depending on the amount of glycolysis agent, the PU glycolysis can be classified in single-phase or split-phase processes. Split-phase glycolysis is mainly obtained using an excess of glycol for the recycling of polyurethanes containing high-molecular-weight polyols. By this method, it is obtained an upper phase consisting

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mostly of the recovered polyols, and a bottom phase including the excess glycol and some by-products of the reaction [6]. Thus, the polyols obtained from split-phase glycolysis have higher purity, and consequently, better properties than those recovered from single-phase processes. The final purity and recovery yield depends mainly on the solubility of the polyols in the glycolysis agent. The development level of glycolysis depends on the glycolysis process considered: single-phase or split-phase. Small-scale commercial glycolysis plants exist in Italy, Austria, Germany, and UK but employing mainly single-phase glycolysis processes with the impurity of recovered polyol that they cause [7]. Split-phase glycolysis processes have been only developed up to pilot plant scale, probably due to the high cost of using the excess of glycols as glycolysis agents. However, recent results have demonstrated the feasibility of performed the split-phase glycolysis of the main flexible PU foams using crude glycerol as glycolysis agent, which is a subproduct from the biodiesel industry and would reduce the costs [4, 8, 9].

To scale-up a chemical process, it is necessary to confirm that the pilot plant scale reproduces geometric similarity criteria and dynamic similarity for laminar flow. According to the literature, in an accurate scaled-up process, it is mandatory that shape factors are maintained constant to ensure a proper geometrical similarity [10].

The goal of this work is to scale-up the glycolysis process from laboratory to pilot plant scale, applying geometric similarity criteria together with dynamic similarity for laminar flow in agitated reactors with the objective of maintaining the same results of recovered polyol properties and mass balance. Hence, a pilot plant installation was designed with geometrically similar equipment to those used for lab scale in a previous work from our research group, obtaining analogous results in terms of recovered polyol properties and mass balance. In addition, a basic design of a split-phase glycolysis industrial plant of flexible PU foams was proposed. Finally, the economic feasibility of such recycling process was confirmed.

Experimental

Materials

Industrial samples of flexible PU foam based on polyether polyol [poly(propylene oxide-block-ethylene oxide)] with molecular weight (Mw) of 3500, functionality with respect to OH groups of 3, polydispersity (PD) of 1.06], and toluene diisocyanate (TDI) were scrapped with an arbitrary diameter ranging from 5 to 25 mm. Crude glycerol was used as glycolysis agent (provided by Biocombustibles de Cuenca S.A.) with a purity of 80%. The 20% of impurities

presented in the crude glycerol corresponds mainly with water, together with a low concentration of salts and free fatty acids. Besides, the obtained results of previous studies have shown that the use of crude glycerol allows to obtain very similar results than those obtained using pure glycerol regarding the recovered products and their quality. Thus, these impurities do not affect significantly to the glycolysis products quality [8]. Stannous octoate was used as catalyst (from Sigma-Aldrich with a purity of 95%). Demineralized water was used as solvent to remove the residual crude glycerol present in the recovered polyol.

Experimental setup

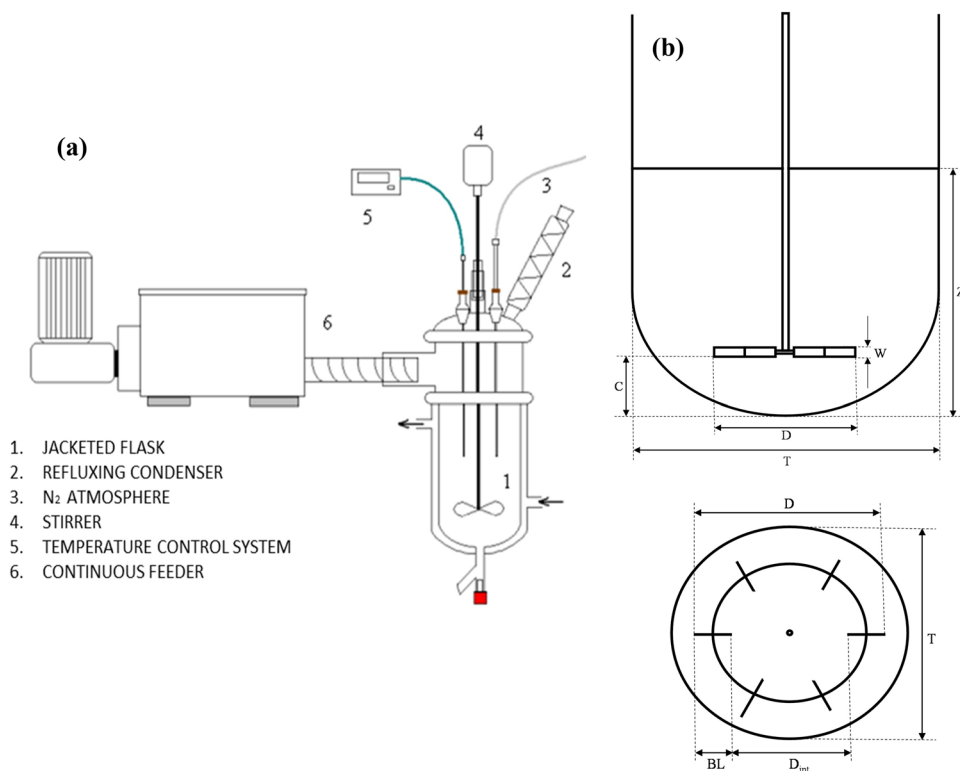
The glycolysis process reactions at lab and pilot plant scales were carried out in jacketed tank reactors with stirring rate and temperature control. The experimental setups had also a condenser with flowing cold water to avoid vapors scape. Foam waste scraps were fed by means of an Archimedean screw from a closed hopper helped by a nitrogen stream. Nitrogen was also constantly fed to the reactor to work under inert atmosphere. The schematic diagram of the glycolysis system and the detail of reactor and stirrer dimensions are shown in Fig. 1.

Characterization

Molecular weight and product composition determination by gel permeation chromatography (GPC)

To carry out the determination of the molecular weight and the distribution of molecular weights, a Viscotek GPCmax VE-2001 TDA302 Detectors chromatograph was employed with two peristaltic pumps, automatic injection system, rack for 120 vials, electric oven, two columns Water Styragel Column HR2 (pore size 500 Å, molecular weight 0–100 g mol⁻¹) and HR0.5 (pore size 50 Å, molecular weight 500–20,000 g mol⁻¹), and triple detection, consisting of a LALS (low angle light scattering) detector, a RALS (right angle light scattering) detector, and a viscosity detector. The GPC equipment was supplied by Malvern Products Iesmat company (Madrid, Spain) and the columns by Waters Cromatografía, S.A. (Barcelona, Spain). The set of columns allowed the detection of molecular weights in the range of 10–20,000 g mol⁻¹, suitable for all synthesized products. The equipment was controlled by a computer system provided with the OmniSEC 4.5.6 program that recorded and analyzed the results. The test conditions were a temperature of 40 °C, a flow rate of 1 mL min⁻¹, a sample concentration of 10 mg mL⁻¹ and an injection volume of 50 µL. The area of a peak was related to the concentration of the substance in the sample by means of the response factors, which were obtained by calibrating with samples of pure substances.

Fig. 1 **a** Schematic diagram of the glycolysis experimental setup. **b** Detail of reactor and stirrer dimensions (T : tank diameter, D : stirrer diameter, C : distance from tank bottom to stirrer, W : impeller blades width; Z : depth of fluid in tank; BL : impeller blade length; D_{int} : stirrer internal diameter)



Structural studies using Fourier transform infrared spectroscopy (FTIR)

The characterization of the obtained products was carried out using a Varian 640-IR FTIR spectrophotometer in the range of $4000\text{--}400\text{ cm}^{-1}$, 8.0 cm^{-1} resolution and 16 scans, with a program called Varian Resolution Pro Software, version 5.0.

Economic assessment

The industrial glycolysis process will be profitable if the income that generates presents a superior value to the required investment to get it. Moreover, that profitability has to be greater than the profitability that provides an alternative employment of the money or than the interest rate to be paid for the borrowed money to carry out the project.

For the economic assessment, the money chronological value must be taken into consideration. For that purpose, three indicators were employed: The Net Price Value (NPV), the Internal Rate of Return (IRR), and the payback period. These three parameters are based in the cash flows analysis with the time. The NPV emerges of summing the cash flows that take place during the investment time horizon, including the initial investment, considering the Weighted Average Cost of Capital (WACC) as discount rate (K). Hence, NPV value is defined as shown in the following equation:

$$NPV_{(K)} = \sum_{i=0}^n \frac{F_i}{(1 + K)^i}, \tag{1}$$

where F_i is cash flow generated during the year i .

If the NPV is higher than zero, the investment would be profitable and the higher the value of NPV is, the higher the economic interest of the project implementation.

The Internal Rate of Return (IRR) is the internal rate that makes the NPV equal to zero (Eq. 2). The higher is the IRR value with respect to the Weighted Average Cost of Capital (WACC), the greater the project profitability:

$$\sum_{i=0}^n \frac{F_i}{(1 + IRR)^i} = 0. \tag{2}$$

On the other hand, the payback period is defined as the length of time required to recover the cost of an investment. An investment with a shorter payback period is considered to be better, since the investor’s initial outlay is at risk for a shorter period of time.

The calculation of these three indicators, and therefore, the profitability study, requires an investment detailed analysis (capital investment, sales, and costs) and the determination of the cash flows. All these details have been calculated and are shown in “[Economic assessment of the split-phase glycolysis plant](#)”.

Results

Experimental setup scale-up from lab scale to pilot plant

To achieve a proper scale-up of the glycolysis process, it was necessary to take into account several points: shape factors requirements in the tank reactor, agitator dimensions, required agitator head power, and rotational speed.

Shape factors in the tank reactor and agitator dimensions

In an accurate scaled-up process, it is mandatory that shape factors are maintained constant to ensure a proper geometrical similarity [10]. In the case of an agitated tank reactor, the shape factors (S1, S2, and S3) are those indicated in the following equations:

$$S1 = D/T \quad (3)$$

$$S2 = Z/T \quad (4)$$

$$S3 = C/T \quad (5)$$

where T is the tank diameter, D is the impeller diameter, C is the distance from tank bottom to impeller center line, and Z is the depth of fluid in tank.

The optimal values of these three shape factors are between 0.7 and 1.0 for the D/T in laminar flow, Z/T can range from 0.5 to 1.0, and C/T must be in the range between 0.1 and 0.4 [10, 11].

For the laboratory device, the value of these parameters were a T of 100 mm and a D of 80 mm and the impeller was placed at 33.3 mm from the bottom, to have a ratio C/T of 1/3 [11]. The depth of fluid in tank (Z) was calculated considering the split-phase glycolysis process at lab scale, obtaining a value of 80.7 mm. Thus, according to these dimensions, the laboratory device shape factors are in the optimal range for an agitated tank reactor ($D/T=0.8$, $Z/T=0.807$, and $C/T=0.333$), also demonstrating its suitability to be scaled-up. In this work, the pilot plant was designed for a production 10 times higher than the lab scale one.

The pilot plant dimensions were calculated maintaining constant the shape factors from the small scale to the large one to preserve the geometrical similarity, which affects strongly to the fluid dynamic of the process. It is also a crucial task to design properly the impeller dimensions, maintaining the geometric similarity with respect to the laboratory scale device, which was a Rushton turbine stirrer with six vertical blades. The main reactor vessel and impeller dimensions of the laboratory and pilot plant scales are shown in Table 1.

Table 1 Main reactor and impeller dimensions of the laboratory and pilot plant scales

	Dimensions (mm)	
	Laboratory scale	Pilot plant scale
Tank diameter, T	100	201
Distance from tank bottom to impeller center line, C	33.3	67
Depth of fluid in tank, Z	80.7	162
Impeller diameter, D	80	161
Impeller internal diameter, D_{int}	50	101
Impeller blade length, BL	15	30
Impeller blades width, W	20	40

Power calculation of the impeller

The power consumed by an agitator, P , can be expressed as a function of the following variables:

$$P = f(D, g, \mu, \rho, N),$$

where g is the gravitational acceleration, μ the fluid viscosity with a value of $1.2065 \text{ kg s}^{-1} \text{ m}^{-1}$, ρ is the fluid density and its value is 1183 kg m^{-3} and N is the impeller rotational speed.

By means of the dimensional analysis, the variables can be grouped in several dimensionless modules shown in the following equation and the next following equation:

$$\frac{P}{D^5 \cdot N^3 \cdot \rho} = K' \cdot \left(\frac{D^2 \cdot N \cdot \rho}{\mu} \right)^b \cdot \left(\frac{D \cdot N^2}{g} \right)^c \quad (6)$$

The number of the dimensionless modules are

$$P_0 = K' \cdot \text{Re}^b \cdot \text{Fr}^c, \quad (7)$$

where P_0 is the Power number, Re is the Reynolds number, and Fr is the Froude number.

In this analysis, Power number is analogous to a friction factor, Reynolds number can be seen as a relation between the inertia forces and the viscous ones, and the Froude number is a relation between the inertia forces and the gravitational one [12, 13].

For the studied glycolysis process, the Reynolds number had a value of 31.4. For Re values lower than 300, the Froude number has not any significative effect. Hence, Eq. 5 can be written as the following equation:

$$P_0 = K' \cdot \text{Re}^b, \quad (8)$$

where b has a value of -1 .

Therefore, the power consumed by the device can be obtained as indicated in the following equation:

$$P = K' \cdot \mu \cdot D^3 \cdot N^2 \tag{9}$$

By means of employing the turbine power correlation for baffled vessels, which represents the Power number versus the Reynolds one [10], it is possible to obtain the Power number taking into account the kind of impeller employed (6 pallets Rushton type agitator), the Reynolds number, and the figure presented in the referenced research from Padron [14].

A value of $P_0 = 4.25$ was obtained and, with it, K' was calculated, having a value of 133.4.

Finally, the required agitation power for the laboratory scale can be obtained from Eq. 9:

$$\begin{aligned} P &= K' \cdot \mu \cdot D^3 \cdot N^2 \\ &= 133.4 \cdot 1.2065 \frac{\text{kg}}{\text{s m}} \cdot 0.08^3 \text{m}^3 \cdot 5^2 \frac{\text{rev}^2}{\text{s}^2} \\ &= 2.06 \frac{\text{kg m}^2}{\text{s}^3} = 2.06 \text{W}. \end{aligned}$$

To ensure a proper scale-up process, it is necessary to calculate the power consumption for the pilot plant scale by means of using the following equation:

$$P_2 = V_2 \cdot \frac{P_1}{V_1} \cdot \left(\frac{T_2}{T_1} \right)^s \tag{10}$$

where P_i are the required agitation powers ($P_1 = 2.06$ W for lab scale), V_i the tanks volumes ($V_1 = 1$ L, $V_2 = 10$ L), T_i the tank diameters ($T_1 = 100$ mm, $T_2 = 201$ mm) and s is the exponent of the laminar flow scaled-up process ($s = -2$ for equal tip speed criterion) [10]. Therefore, the agitation power required for the pilot plant scale was of 5.1 W.

Impeller rotational speed

Another crucial task to ensure a successful scale-up is to determine the impeller rotational speed at pilot plant scale by means of employing Eq. 9, the constant K' , and the agitation power values estimated previously, with a value of 133.4 and 5.1 W, respectively [10]:

$$\begin{aligned} N &= \left(\frac{P}{K' \cdot \mu \cdot D^3} \right)^{\frac{1}{2}} = \left(\frac{5.1 \frac{\text{kg m}^2}{\text{s}^3}}{133.4 \cdot 1.2065 \frac{\text{kg}}{\text{s m}} \cdot 0.1608^3 \text{m}^3} \right)^{\frac{1}{2}} \\ &= 2.76 \text{rps} \cdot \frac{60 \text{s}}{1 \text{min}} = 165.63 \text{rpm}. \end{aligned}$$

Glycolysis process results at pilot plant scale

To confirm the suitability of the selected criteria for the scale-up process, the glycolysis process experiment of PU foam wastes was carried out at pilot plant scale at the same conditions than at lab scale. This is at 190 °C and inert atmosphere

Table 2 Glycolysis reaction recipe at pilot plant scale

	Mass (g)
PU foam	3000
Crude glycerol	4441.5
Stannous octoate	58.5

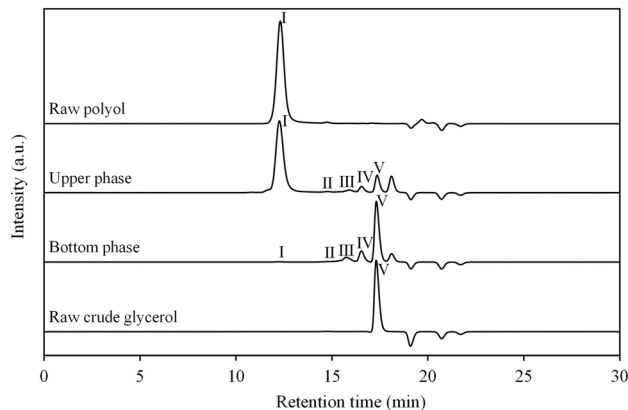


Fig. 2 GPC chromatograms of the glycolysis product at the end of reaction in comparison with GPC chromatograms of raw flexible polyether polyol and raw crude glycerol. Peak I=flexible polyether polyol; Peak II, III and IV=reaction by-products; Peak V=crude glycerol

and with a crude glycerol to PU foam waste ratio of 1:1.5 and a catalyst concentration of 1.3 wt% [2, 8, 15–17]. The reaction recipe at pilot plant scale appears in Table 2.

As expected, a split-phase product was obtained from the glycolysis process at pilot plant scale. Figure 2 shows the GPC analysis of the upper and bottom phases from the pilot plant experiment compared to those from the raw polyether polyol and crude glycerol.

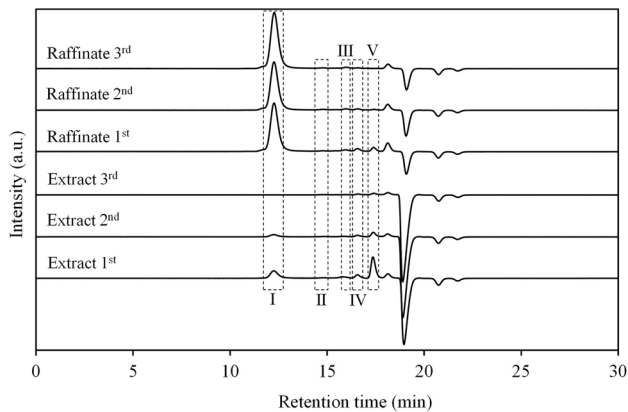
From these results, the split-phase glycolysis process has been demonstrated at a pilot plant scale, using the same conditions than at laboratory scale. Since an upper phase has been obtained, consisting mainly of recovered polyol (Peak I) and a low concentration of reaction by-products (Peak II–IV) and glycolysis agent solubilized in this phase (Peak V). On other hand, the bottom phase presented the excess of glycolysis agent used in the reaction together with the reaction by-products and a low concentration of recovered polyol solubilized in this phase.

To quantify the quality of the glycolyzed product, the concentration of the different compound in wt% of each phase were obtained by means of GPC analysis [18]. The results are shown in Table 3.

Next, to improve the purity of the recovered polyol, the upper phase was washed with water, solubilizing the reaction by-products and the glycolysis agent in water, and achieving an upper phase with a higher concentration of recovered polyol, process carried out in previous studies [2]. Three

Table 3 Concentration in wt% of the different compounds in the upper and bottom phases at the end time of the reaction

	Composition (wt%)	
	Upper phase	Bottom phase
Recovered polyol	81.0	0.4
Reaction by-products	7.7	25.1
Crude glycerol	11.3	74.5

**Fig. 3** GPC chromatograms of the raffinates and extracts of extraction process. Peak I=flexible polyether polyol; Peak II, III and IV= reaction by-products; Peak V= crude glycerol

consecutive extractions were performed. Figure 3 shows the GPC chromatograms of each of the extractions, both raffinates and extracts.

An upper phase has been obtained with higher purity in recovered polyol and lower concentration of by-products and glycolysis agent. Besides, analogously to the calculations of the upper and bottom phases, the concentration of each compound in the raffinate phases has been determined [18]. These results are shown in Table 4.

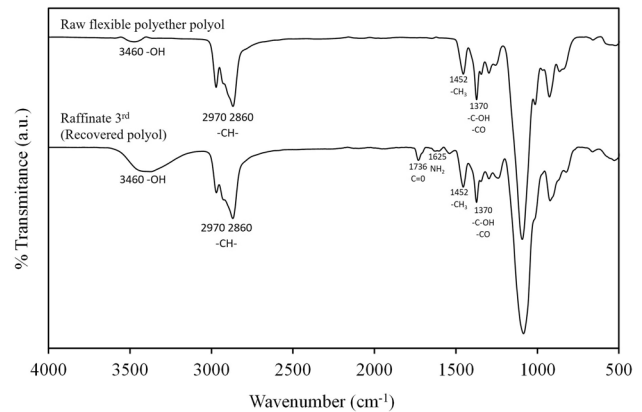
From these results, it can be highlighted that a product with a purity higher than 96% in recovered polyol has been obtained, capable of replacing the pure polyol in the synthesis of new polyurethane foams.

Furthermore, the polyol recovery yield was estimated from the recipe of glycolized foam (62 g of polyol per 100 g of foam), the purity of recovered polyol (96.2%wt) and the mass balances in which were obtained 1820 g of recovered polyol for every 3000 g of glycolized foam, resulting in a value of 93.9%. According to these data, the polyol recovery yield was

$$\frac{1820 \text{ g recovered polyol}}{3000 \text{ g glycolized foam}} \cdot \frac{96.2 \text{ g polyol}}{100 \text{ g recovered polyol}} \cdot \frac{100 \text{ g glycolized foam}}{62 \text{ g polyol}} = 93.9\%$$

Table 4 Concentration of the different compounds in the raffinate phases

	Composition (wt%)		
	Raffinate 1°	Raffinate 2°	Raffinate 3°
Recovered polyol	89.2	94.7	96.2
Reaction by-products	6.6	4.5	3.5
Crude glycerol	4.2	0.8	0.3

**Fig. 4** FTIR spectra of the raw polyol and the raffinate or recovered polyol

Finally, to complete the characterization of the obtained products, FTIR analyses were carried out. Figure 4 shows the spectra of the upper and bottom phases together with the last raffinate.

The FTIR spectra of raw polyol and raffinate presented the typical functional groups of a polyol, showing the signals corresponding to the hydroxyl groups OH at 3460 cm^{-1} , CO groups associated with hydroxyl groups at $1000\text{--}1400 \text{ cm}^{-1}$, CH bonds of the aliphatic carbons of the chain polyol at $2800\text{--}3000 \text{ cm}^{-1}$, and methylene groups at 1452 cm^{-1} [19, 20]. Besides, the FTIR spectra of the raffinate presented two additional signals associated to the reaction by-products of the glycolysis process. These signals correspond to the C=O groups (1736 cm^{-1}) produced by the transesterification carbamates and the amine groups produced by NH_2 signal (1625 cm^{-1}) from products similar to toluenediamine [8].

From the results shown in this section, it can be concluded that the split-phase glycolysis experiments carried out at the pilot plant scale confirmed the technical feasibility of the recycling process at larger scales, since analogous results in terms of recovered polyol properties was obtained, comparable with previous studies carried out at laboratory scale [2, 8, 15–17].

Economic assessment of the split-phase glycolysis plant

Once confirmed the success in the scaling-up of the process, the design and the economic assessment of a split-phase glycolysis industrial plant was addressed.

Glycolysis plant capacity

With the aim of determining the capacity of the glycolysis industrial plant, it is an essential issue to know the PU production and the quantity of waste generated nowadays in our country. To get this goal, first, the Spanish PU production was estimated from that of Western Europe (WE).

In 2013, the PU production in WE was 3.35 MTm with an approximately annual growth of annual growth between 2013 and 2018 of about 1.9% [21]. Thus, the estimated production in 2020 would be close to 3.82 MTm. Once known this value, and comparing the production of PU foams between WE and Spain in 2004 (2.5 MTm and 178,259 Tm, respectively), it was estimated the production of PU foams in Spain in 2020, obtaining a value of 272,506 Tm [22, 23].

The designed glycolysis process has been mainly developed for flexible PU foams, which involve the 36% of the PU products [21]. Thus, the Spanish production of flexible PU foams in 2020 will be about 98,102 Tm.

As the plant is design for the PU foam waste generation, the calculated Spanish production of flexible PU foams has to be related to the waste generation. It is known that the PU waste generation in WE in 2004 was 1.5 MTm, considering not only post-consumer products but also scrap from slab stock manufacturing [22]. Thus, the PU waste was about the 60% of the production. Applying this percentage to the Spanish production of flexible PU foams, the PU waste generation in 2020 in Spain will be of about 58,861 Tm.

Moreover, it is estimated that more than the half of the total polyurethane waste is not suitable for collection and subsequent recycling, because of small volume and/or wide distribution. This waste is best recovered by combustion (incineration with energy recovery) together with municipal solid waste [22]. Therefore, considering that just about the 45% of flexible PU foam waste is susceptible of being recycled in Spain, nowadays it would be a quantity close to 26,500 Tm and the plant for this study is going to be designed to recycle about 1% of this waste production. Thus, the plant capacity is fixed on 270 Tm.

For this treatment capacity, the mass balance has been carried out for batches, considering that the total time of the split-phase glycolysis process are 12 h per batch, including loading and unloading times and cleaning, taking as reference PU chemical degradation processes operating industrially at present [24]. It was considered 300 working-days/year (600 batches/year) and 3 work shifts of 8 h, giving as

a result 7200 h of production/year. For treating the 270 Tm/year, each batch must treat 450 kg.

Process mass balance and conceptual definition of the split-phase glycolysis process

The consumption of raw materials involved in the recycling by means of split-phase glycolysis of flexible PU foam waste at industrial scale was calculated from the pilot plant scale data and is shown in Fig. 5.

As can be seen in Fig. 5, the industrial recycling process consists of introducing the transesterification agent (glycol and catalyst) into the glycolysis batch reactor (R-1) by means of the gear pumps P-2 and P-3, respectively.

Once achieved the reaction temperature (190 °C), the required quantity of PU foam scraps is added by means of an Archimedean screw (P-1), according to its dissolution. Once finished the glycolysis reaction, the product is discharged using P-4 and it is introduced in the decanter C-4, in which the product is splitted in two phases. Bottom phase is unloading using P-5 to the bottom phase storage tank (C-5). Regarding the upper phase, it is introduced to a liquid–liquid extractor (C-7) by means of P-6, together with wash water. The extract phase is discharged using P-8 and is stored in the wastewater storage tank (C-8). Finally, the recovered polyol is unloading by means of P-9 and is stocked in the recovered polyol storage tank (C-9).

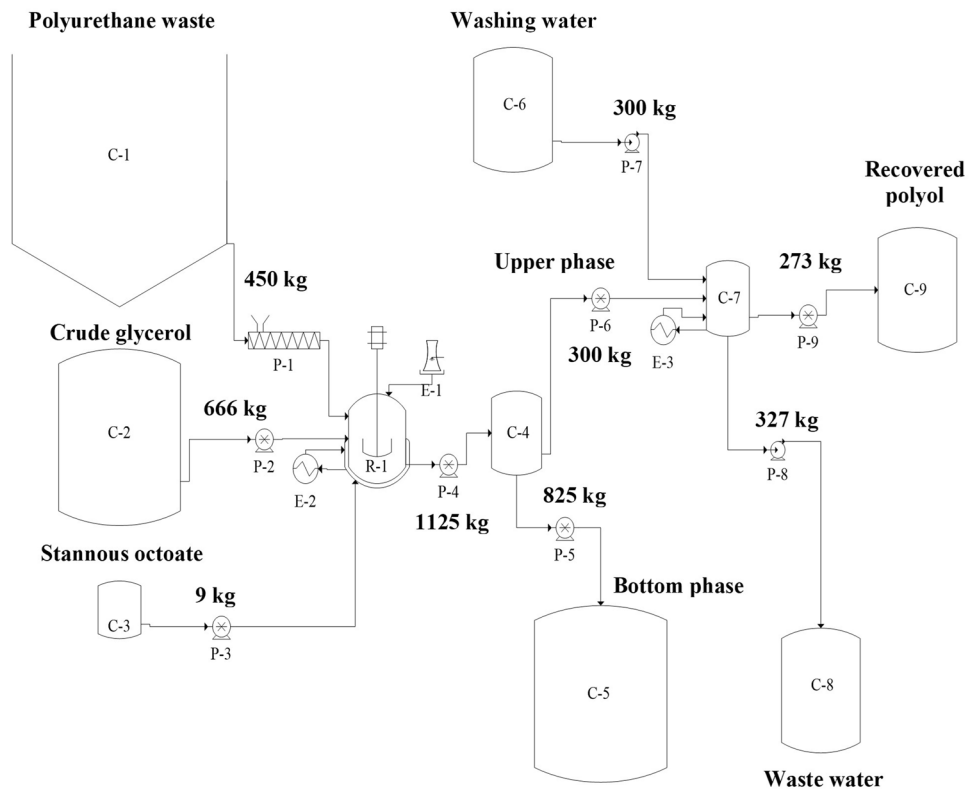
Investment calculation

As indicate above, a process is profitable if the income that generates presents a superior value to the required investment to get it. Thus, in this section, the investment required for the proposed glycolysis plant is determined. The investment has to take into account all the parameters associated to the plant construction: the capital, the sales and the costs.

The capital investment represents both the amount of money which must be supplied for the manufacturing and plant facilities, namely *fixed capital investment*, and that required for the operation of the plant, known as *working capital*.

Fixed capital investment The fixed capital investment is the one employed in the acquisition of production resources that have to be able to remain themselves to make possible the production, hence it cannot be converted into cash and includes grounds, equipment, installation costs, and licenses [25]. It can be calculated by means of the percentages method, in which the fixed capital investment is estimated as percentage of the equipment's costs [26]. Hence, first, it was necessary to know the costs of the equipment, which are gathered in Table 5. The equipment costs were obtained from the webpage www.matche.com or calculated by the method described by Cabra et al. [26, 27].

Fig. 5 Flow diagram of the split-phase glycolysis plant and the mass balance per batch



Using the percentages' method, the fixed capital investment was calculated, and the results are shown in Table 6.

Working capital It is defined as the capital that is easily transformed into cash and, in the case of the glycolysis plant, includes as main items the raw materials stock, the products stock, the spare parts warehouse and cash to attend to payments [26]. In the glycolysis plant, it will be considered as the average stocks of raw materials and products for 15 days, together with the spare parts warehouse and cash to attend to payments, being estimated the last two ones as the 2.5% of the fixed capital investment [26].

Table 7 [28] shows the price of the raw materials (crude glycerol [29] and stannous octoate) together with the estimated sale price of the glycolysis products and the sale price of a raw polyether polyol [30]. The prices of raw materials and products were updated using the Industrial Products Index with a value of 3.6% [31].

According to previous studies [15, 32], the glycolysis upper and bottom phases can be used in the polyurethanes industry for the synthesis of new flexible and rigid PU foams, respectively. Thus, their sale prices were estimated to be the 80% and 40% of the raw polyether polyol price for the upper and the bottom phases, respectively. On the other hand, the price of the PU scraps was considered null since the waste treatment is an offered service to avoid the companies or institutions problems related to the waste treatment.

Table 5 Equipment cost of the split-phase glycolysis plant [25, 26]

Equipment	Cost (€ ₂₀₂₀)
Polyurethane silo (C-1)	48,425
Storage tank (C-2)	19,763
Storage tank (C-3)	1221
Decanter (C-4)	3095
Storage tank (C-5)	22,934
Storage tank (C-6)	13,294
Liquid-liquid extractor (C-7)	2687
Storage tank (C-8)	14,082
Storage tank (C-9)	12,330
Glycolysis reactor (R-1)	4116
Condenser (E-1)	18,852
Heat exchanger (E-2)	15,271
Heat exchanger (E-3)	9504
Archimedeian screw (P-1)	2772
Gear pump (P-2)	2588
Gear pump (P-3)	1386
Gear pump (P-4)	3142
Gear pump (P-5)	2772
Gear pump (P-6)	2218
Centrifugal pump (P-7)	2218
Centrifugal pump (P-8)	2218
Gear pump (P-9)	2126
Total	207,015 €

Table 8 shows the value of the working capital of the split-phase glycolysis plant.

It is interesting to observe that the estimated working capital represents approximately the 10% of the fixed capital investment, which agrees with the typical percentage range recommended in the literature [26].

Sales Sales are defined as the money obtained from the sale of the products generated in the process. Table 9 shows the sales obtained in the split-phase glycolysis process.

In this point, it is interesting to comment that one of the simplest and most used methods to estimate an order of magnitude of the investment is the fixed asset turnover ratio, which relates the sales with the fixed capital investment, according to the following equation [26]:

$$\text{Fixed asset turnover ratio} = \frac{\text{Sales}}{\text{Fixed capital investment}} = \frac{773,506 \text{ €}}{799,558 \text{ €}} = 0.97 \quad (11)$$

Table 6 Fixed capital investment by the equipment cost percentage method [26, 28]

Fixed capital investment	Cost (€ 2020)
1. Equipment cost (E)	207,015
2. Materials (M) 60% E	124,209
Civil work (28% M)	34,779
Piping and infrastructures (45% M)	55,894
Instrumentation (10% M)	12,421
Electricity (10% M)	12,421
Insulation (5% M)	6210
Painting (2% M)	2484
3. Detail engineering (20% (E + M))	66,245
4. Process engineering (10% (E + M))	33,122
5. Construction (50% (E + M))	165,612
6. Construction supervision (10% (E + M))	33,122
ISBL (\sum Previous costs)	629,325
7. Auxiliary services (4% ISBL)	25,173
8. Off-sites (8% ISBL)	50,346
9. Start-up costs (3.5% ISBL)	22,026
10. Contingency (10% (ISBL + 7 + 8 + 9))	72,687
Total cost	799,558 €

Table 7 Price of the raw materials and sale price of the glycolysis product

Chemical	Price/sale price (€ ₂₀₂₀ /Kg)
Crude glycerol	0.12
Stannous octoate	12.22
Raw polyether polyol	2.35
Recovered polyol	1.88
Glycolysis bottom phase	0.94

Table 8 Working capital requirements of the split-phase glycolysis plant

Concept	€ ₂₀₂₀
Stock crude glycerol	2,442
Stock stannous octoate	3,300
Stock raw materials	5,742
Stock recovered polyol	15,402
Stock glycolysis bottom phase	23,273
Stock products	38,675
Spare parts warehouse	19,989
Cash	19,989
Working capital	84,396 €

For the chemical industry, this ratio is approximately 0.95 [26], which agrees with the obtained fixed asset turnover ratio of the split-phase glycolysis plant, demonstrating the consistency of the economic assessment up to this point.

Annual costs They are defined as the conversion in money of the goods and services consumed to produce. Costs include the raw materials, the direct operating labor, the auxiliary services, the directive management, the maintenance, the taxes, the insurances, and the commercial costs [26]. Costs can be divided in *manufacturing costs* (cost of raw materials, direct operating labor, the disposal of waste, and the auxiliary services), and in *general costs* (indirect operating labor, the directive management, the maintenance, the taxes and the insurances). Considering the operating times, three operators and one person for the directive management and other indirect operating labors are needed.

On the other hand, the maintenance was estimated as a 1% of the sum of fixed and working capitals, insurances and taxes, were estimated as a 1% and 0.5% of the fixed capital, respectively. The commercial cost is estimated as a 5% of the manufacturing cost.

Table 10 gathered the main annual costs for the proposed glycolysis plant [33–36].

Cash flow analysis Once the capital investment, the sales and the costs have been determined, the profitability of the project is studied by means of the cash flow analysis, supposing that the capital investment contributions are entirely carried out with own capital (without leverage). Obviously, this is a conservative scenario, since the working capital is normally entirely financed with short-term debt whereas a high percentage of the fixed capital investment is funded with long-term debt, which would increase the project profitability. Besides,

Table 9 Sales of the split-phase glycolysis plant

Product	Mass flow (kg/batch)	Mass flow (Tm/year)	Sale Price (€ ₂₀₂₀ /Tm)	€ ₂₀₂₀ /year
Recovered polyol	273	163.8	1,881	308,048
Glycolysis bottom phase	825	495	940	465,458
Sales				773,506 €

Table 10 Main annual costs

Raw materials (1)			
	Price (€/kg)	Consumption (kg/h)	Cost (€/year)
Crude glycerol	0.12	666	48,848
Stannous octoate	12.22	9	66,010
Direct operating labor (2)			
	Work shifts	Cost (€/operator-year)	Cost (€/year)
Operator	3	25,000	75,000
Waste treatment (3)			
Waste disposal	Mass flow (Tm/year)	Cost (€/Tm)	Cost (€/year)
Extract phase	196,2	511.4 [33]	100,336
Auxiliary services (4)			
	Price (€/kWh)	Consumption (kW/year)	Cost (€/year)
Electricity	0.07015 [34]	360,559	25,293
	Price (€/m ³)	Consumption (m ³ /h)	Cost (€/year)
Nitrogen	2.43 [35]	4611.6	11,206
	Price (€/m ³)	Consumption (m ³ /year)	Cost (€/year)
Wash water	1.2688 [36]	180	228
Total direct manufacturing costs [(1) + (2) + (3) + (4)] (€/year)			326,923
Indirect manufacturing costs		Estimation [26]	Costs (€/year)
Directive management/technologist		–	45,000
Maintenance		1% (Fixed capital investment + working capital)	8840
Taxes		0.5% (Fixed capital investment)	3998
Insurances		1% (Fixed capital investment)	7996
Total indirect manufacturing costs (€/year)			65,834
General costs (commercial costs) (5% direct + indirect costs) (€/year)			19,635
Total annual costs (direct and indirect costs + general costs) (€/year)			412,392 €

we have used a WACC value of 8%, since it has been the recent value considered as discount rate by several chemical industries [37]. The inflation value considered corresponds with the Spanish inflation average value for the last 4 years (1.16%) [38]. The value of inflation has been employed to update the value of annual sales and costs for each year. In addition, it is crucial to fix a project time horizon, which for chemical plants of the dimensions of the proposed plant is between 10 and 15 years. In the present work, a time horizon of 15 years has

been considered, proposing to carry out the whole investment in the zero year (without leverage) with a linear amortization in 10 years (10%), and in last year, the working capital is recovered increasing the cash flow of the year 15. What is more, focusing on a conservative scenario, it has been considered that the level of production during the first operation year of the industrial glycolysis plant is the 50% of the glycolysis plant total capacity, the 80% during the second year and the 100% during the rest of the time horizon considered. Finally,

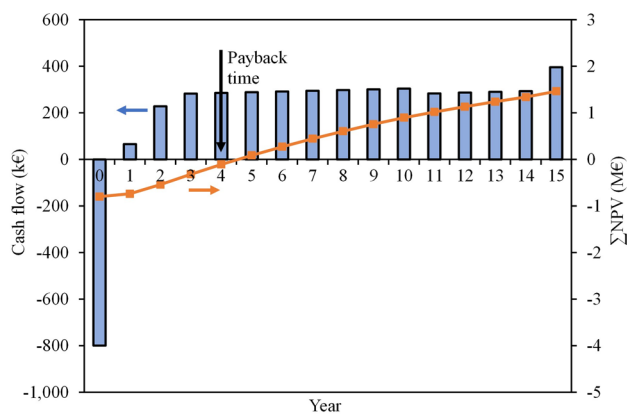


Fig. 6 Annual cash flow of the split-phase glycolysis plant

it is necessary to consider a tax rate, that in this work has been estimated of 30.0%. Considering all these data, the cash flow was calculated and is shown in Fig. 6.

The cash flow analysis demonstrates the project economic viability, since it has been obtained a NPV of 1,464,555€ with an IRR of 27.99%, according to Eqs. (1) and (2), respectively, and a payback time between 4 and 5 years. It has to be noted that a positive NPV has been obtained and that the IRR is considerably higher than the considered $K = 8\%$, demonstrating the profitability of the split-phase glycolysis plant investment.

Economic sensitivity analyses

The use of the percentages method for the investment calculation implies some uncertainty level. Hence, it was considered to perform an investment sensitive study considering an investment fluctuation of $\pm 20\%$. The results are shown in Fig. 7.

It can be appreciated in Fig. 7 that, logically, the higher the required investment the lower the NPV and the IRR obtained. Nevertheless, it has been demonstrated that, even in the worst scenario (+20% investment), the split-phase glycolysis plant investment would provide considerably higher profitability than an alternative employment of the money, ensuring the profitability of the designed recycling plant.

Besides, since the price of polyol can fluctuate, an economic sensitivity analysis has been carried out considering a fluctuation of $\pm 20\%$. The results are shown in Fig. 8.

From the results obtained, it can be confirmed that the process is economically profitable, even in the worst scenario, when the price of polyol is reduced by 20. Besides, it has been verified that if the price of polyol increased by 20%, it would imply an increase in the NPV and the IRR of 52 and 35%, respectively. Thus, after considering possible fluctuations in the price of polyol, it can be also stated that the process is quite robust from the economical point of view.

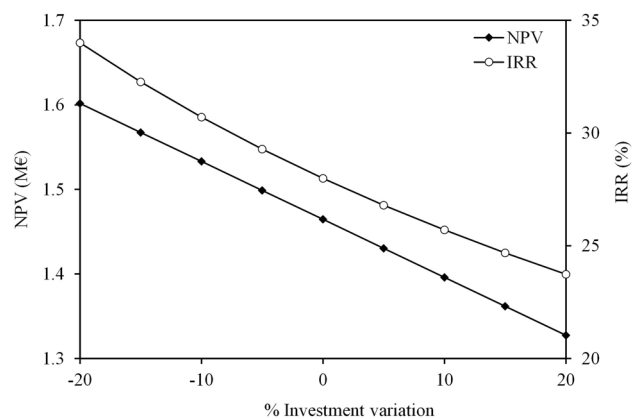


Fig. 7 Investment sensitivity study

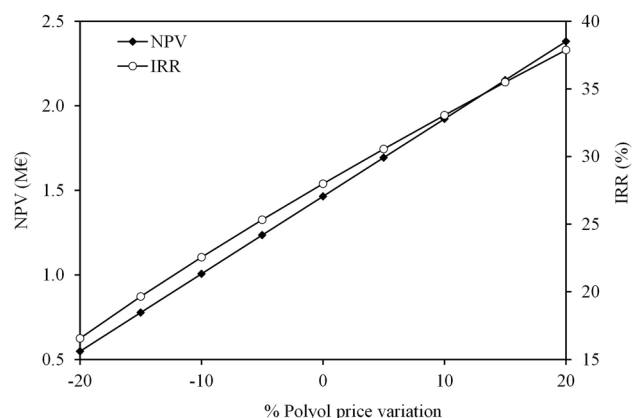


Fig. 8 Polyol price sensitivity study

Conclusions

The scaled-up feasibility of the split-phase glycolysis process has been confirmed by means of designing a pilot plant installation similar geometrically to that used for lab scale for a production 10 times higher than the lab scale one; obtaining analogous results in terms of recovered polyol properties and mass balance. With the aim of performing this task, geometric similarity criteria have been applied together with dynamic similarity for laminar flow in agitated tank reactors.

Moreover, taking as starting point the results obtained in the pilot plant scale, it has been demonstrated the economic viability of a split-phase glycolysis industrial plant with a capacity for treating 270 Tm per year of flexible PU foams scraps. From this study, a Net Present Value (NPV) of 1,464,555€, an Internal Rate of Return (IRR) of 27.99%, and a payback time between 4 and 5 years have been obtained. Furthermore, the projected Income Statement shows that the glycolysis plant turns a profit from the first year of operation.

In addition, the investment sensitive study and the polyol price sensitive study have demonstrated that even in the worst scenario considered (+ 20% investment and – 20% polyol price), the split-phase glycolysis plant investment would provide a high profitability.

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