

b – CAROTENE PRECIPITATION BY PGA PROCESS. SCALE UP STUDY

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ABSTRACT

A pilot plant for the precipitation of solids from organic solutions by addition of SC CO₂ as antisolvent has been used to study the scale up of the precipitation process of β carotene from dichloromethane solutions. Experimental data were obtained for β carotene at different operating conditions, such as pressure, β carotene concentration on the organic solution, organic solution mass rate, antisolvent mass rate, and also the mass ratio SC – CO₂ / solution. The influence of the operation variables and the mixing system on the particle size of the obtained solid has been studied in a previous paper. In this work from some new experimental data it is presented a complete flow diagram for the production of β carotene at industrial scale. The following data are obtained: Plant Capacity: 7 200 kg β carotene /year; Precipitation conditions: 9.0 MPa and 35 °C; SC CO₂ mass rate: 12 680 kg/h; solution mass rate: 2 024 kg/h.

INTRODUCTION

PGA process has been successfully applied to β carotene crystallization from dichloromethane solutions using CO₂ as antisolvent. In a previous work, it was studied the crystallization of β carotene by a batch GAS process. Results show GAS process yields trans isomer with high efficiency. Crystals smaller than 1 μ m are obtained at temperature of 298 K and 5.8 MPa from 1 g/L solutions of β carotene, but it was observed a high degree of agglomeration [1]. The mean concentration of CO₂ is 90 %_w therefore, the volume of the crystallizer has to be of 290 L to obtain 1 g of β carotene. So a continuous antisolvent process could be the alternative for the industrial application of the β carotene crystallization. It has also been studied the influence of several variables in the continuous process [2]. In that work, it was observed that precipitation form varies with the mixing system and the mass flow rate. It was also observed that lower precipitation pressures produce more agglomerated crystals.

GAS and PGA process has been applied to other natural products, such as proteins, pharmaceutical products, and inorganic compound precipitation. See Reverchon [3] and Perrut [4] for reviews.

The aim of this work is to study the mass and energy balances for an industrial plant for the precipitation of β carotene by continuous antisolvent precipitation with SC CO₂.

EXPERIMENTAL

2.1. Experimental apparatus

A continuous pilot plant for continuous GAS process is employed for this project. The flow diagram of the installation is presented in figure 1.

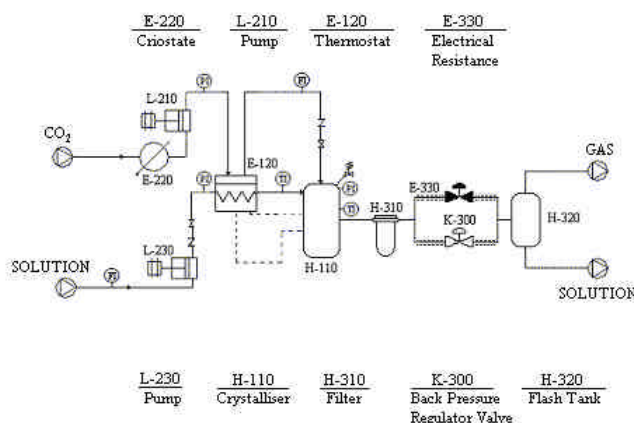


Figure 1. Flow diagram of the GAS plant

The equipment used are two diaphragm pumps (Dosapro, Spain), one for the CO₂ (L – 210) and the other for the solution (L-230). An isolated and jacketed AISI 316 stainless steel crystallizer (H – 110) with a metal porous frit at the exit. There is also an external stainless steel filter (H- 310) from Headline filters (UK). This filter presents a screen size of 1 μ m. A separation flask (H – 320) to achieve the separation of the solvent and the CO₂, by pressure release. The other elements are flow, pressure and temperature meters and heat exchangers to achieve the operating conditions for the pilot plant. The diffuser system is an atomizing nozzle (1/4J SU 22) from Spraying Systems Co. (Spain). For further details, see Cocero [2].

2.2. Reagents and materials

Crystalline β carotene with a minimum purity of 99.5% was obtained from *ANTIBIOTICOS S.A.* (Spain). Dichloromethane (99.5%) was provided by *PANREAC QUIMICA, S.A.* (Spain). Carbon dioxide (99.99%) was supplied by *CARBUROS METÁLICOS, S.A.* (Spain).

2.3. Analytical procedures

Concentration of solutions and purity of crystals was analyzed by HPLC, using a Waters 996 Photodiode Array Detector as detection system and a Waters 600 Controller as pump. The eluent was detected at 450 nm for ethyl acetate and 445 nm for dichloromethane. The column was C-18, Nova-Pack, 60A, 4 μ m, 3.9 \times 300 mm. As mobile phase 90/10 methanol–chloroform, 1 ml/min isocratic, at 293 K was used.

A JEOL JSM-820 Scanning Electron Microscope (SEM) has been used for obtaining the photomicrographs of the crystals.

RESULTS AND CONCLUSIONS

3.1. Experimental results.

In previous essays, it was studied the influence of the pressure, temperature, initial solution concentration, % $w\text{CO}_2$, mixing system and solution and SC CO_2 rates. From them, it is stated that mixing systems influences particle size, PSD and agglomeration of the particles. This influence is due to the change in the precipitation mechanism from growing to nucleation. It depends on the mixing degree achieved between solvent and antisolvent. If the mixing is poor a big droplet of solution is formed and then the SC CO_2 has to diffuse to cause the precipitation of the solute (growth mechanism). If the streams are mixed properly, a mist is formed and then precipitation happens due to the nucleation mechanism. A similar behavior is described for Lobenzarit solution by Amaro Gonzalez et al [5]. So to let nucleation mechanism control the precipitation and get smaller particles, it is used a commercial diffuser from Spraying Systems Co. (Spain) ($\frac{1}{4}$ J SU 22).

In this work it is studied the influence of the pressure on both particle size and yield of the process. In figure 2, it is presented the β carotene obtained at 90 bar. As it can be seen, needle like crystals are obtained with sizes ranging from 50 to 200 μm . In figure 3, it is presented the β carotene obtained at 150 bar. In this run, β carotene ranging from 10 to 200 μm is obtained. It was observed that particle size decreases with pressure and also precipitation yield did so. The decrease in the particle size can be associated with the kinetic of the mixing process. Although both points are in one phase condition inside the crystallizer, as pressure increases, the kinetic of the mixing is also increased because of the solvent power of the SC CO_2 to the solvent is increased. This decrease in the particle size may be due to the enhancement of the solvent power of CO_2 with increasing pressure. Reverchon et al. [6] describe the same behavior for the precipitation of tetracycline from N-methyl-2-pyrrolidone solutions.

	P	T	MCO2	Mdis	CO2/SOL	Cfeed	Cres_sol	Rdto	top
	bar	°C	kg/h	kg/h	%	ppm	ppm	%	min
BDM35N2	90.1	35.1	3.18	0.508	86.4%	698	41	97.1%	105
BDM35N3	89.9	34.1	2.84	0.482	86.6%	623	47	97.6%	75
BDM35N4	89.9	34.3	3.10	0.487	86.5%	632	43	96.5%	110
BDM35N5	79.9	33.4	3.12	0.492	86.3%	681	185	95.8%	50
BDM35N6	150.3	33.2	3.08	0.473	86.3%	662	248	90.7%	70

Table 1. Experimental runs and operating conditions.

3.2. Scale up study.

For the scaling up process, increasing the pressure presents the drawback that the precipitation yield decreases but the advantage that smaller particles are obtained.

From the experimental data, it is studied the scale up of the plant to the production of 7 200 kg/year of β carotene. So, keeping the experimental conditions of pressure and temperature the resulting flow diagram is presented in figure 4. The resulting mass and energy balances are presented in table 2. As it can be seen, the plant will work with 12 680 kg CO_2 /h and 2 024 kg dichloromethane /h, at 90 bar and 35 °C.

The main operations are the precipitation and the recovery of both solvent and antisolvent. The precipitation conditions are set in order to get a maximum precipitation yield. So, the design is made for 90 bar and 35 °C, in order to achieve one phase precipitation inside the precipitator (according to Soave-Redlich-Kwong EOS prediction, complete miscibility is achieved at about 83 bar). At the same time, it is also important to keep the same flow rate ratio between CO₂ and dichloromethane as in the experimental plant to be able to control the antisolvent concentration in the precipitation vessel as well as the velocity ratio in the mixing system.



Figure 2. SEM photograph from BDM35N2, at 90 bar ($r = 100$ mm)



Figure 3. SEM photograph from BDM35N6, at 150 bar ($r = 100$ mm)

The flow diagram is divided in two different sections, one for the precipitation of the β carotene and another one for the recuperation of both solvent and antisolvent. In the first section, there are two vessels, one for the antisolvent and another one for the solution. In order to reduce recompression costs of the recirculation streams, these vessels are working at 50 bar (the same pressure as the separating vessel). From vessel D – 101 solution is pumped to the precipitator pressure by a membrane pump (L – 102), and heated to the precipitator temperature by means of heater (E – 103). From vessel D – 201 antisolvent is pumped to the precipitator pressure by a membrane pump (L – 202), and heated to the precipitator temperature with a heater (E – 203).

Once both streams are at the pressure and temperature of the precipitator, they are mixed inside the precipitator to achieve the precipitation of the β carotene. Then, a mixture exits the precipitator to enter a filter where the β carotene is recovered. After the filter, the mixture is decompressed to the storing pressure. After this decompression, the stream is heated to achieve the operation temperature of the distillation tower where the solvent is recovered with a low concentration of antisolvent. This solvent stream (with an antisolvent concentration of 7 %_w) is recirculated to the solvent storage vessel (D - 101). The antisolvent stream is also recirculated to its storage vessel (D - 201). From this stream a purge stream is taken to avoid the accumulation of undesirable compounds in the system.

In the precipitation vessel, it is very important to achieve a suitable mixing degree to be able to control both particle size and precipitation yield. In order to get this, two alternatives can be used. The first alternative is a commercial mixing system like the one employed in this research (from Spraying Systems Co, SPAIN), where both fluids are mixed just before entering the vessel. The second alternative is the usage of a steel frit, in which the pore size is controlled to achieve a suitable droplet size of the liquid solution.

N° Corriente	1	2	3	4	5	6	7	8	9	10	N° Corriente
P (bar)	49.0	89.0	89.0	49.0	49.0	89.0	89.0	89.0		89.0	P (bar)
T (K)	323.15	325.55	308.15	288.65	263.15	266.75	308.15	317.35		317.35	T (K)
Entalpía (KJ/h)	-4.30E+06	-4.30E+06	-4.34E+06	-1.15E+08	-1.18E+08	-1.18E+08	-1.16E+08	-1.20E+08		-1.20E+08	Entalpía (KJ/h)
CO2 (kg/h)	154	154	154	12680	12680	12680	12680	12834		12834	CO2 (kg/h)
DCM (kg/h)	2024	2024	2024	47	47	47	47	2071		2071	DCM (kg/h)
b car (kg/h)	1.06	1.06	1.06	1.06	1.06	1.06	1.06	1.06	1.00	0.06	b car (kg/h)
TOTAL(kg/h)	2179	2179	2179	12728	12728	12728	12728	14906		14906	TOTAL(kg/h)
CO₂ (kmol/h)	3.49	3.49	3.49	288.12	288.12	288.12	288.12	291.62		291.62	CO₂ (kmol/h)
DCM(kmol/h)	23.83	23.83	23.83	0.55	0.55	0.55	0.55	24.38		24.38	DCM(kmol/h)
TOTAL (kmol/h)	27.33	27.33	27.33	288.67	288.67	288.67	288.67	316.00		316.00	TOTAL (kmol/h)

N° Corriente	11	12	13	14	15	16	17	18	19	20	N° Corriente
P (bar)	50.0	50.0	49.3	50.8	49.3	49.3	49.0		49.0	50.8	P (bar)
T (K)	296.05	303.15	289.65	445.45	289.65	289.65	285.15		285.15	323.15	T (K)
Entalpía (KJ/h)	-1.20E+08	-1.19E+08	-1.14E+08	-3.94E+06	-5.72E+06	-1.09E+08	-5.84E+06		-4.19E+03	-4.30E+06	Entalpía (KJ/h)
CO2 (kg/h)	12834	12834	12680	154	634	12046	634		0	154	CO2 (kg/h)
DCM (kg/h)	2071	2071	49	2022	2	47	0		3	2022	DCM (kg/h)
b car (kg/h)	0.06	0.06	0.00	0.06	0.00	0.00	0.00	1.00	0.00	0.06	b car (kg/h)
TOTAL(kg/h)	14905	14905	12729	2176	636	12093	634		3	2176	TOTAL(kg/h)
CO₂ (kmol/h)	291.62	291.62	288.12	3.49	14.41	273.72	14.41		0.00	3.49	CO₂ (kmol/h)
DCM(kmol/h)	24.38	24.38	0.58	23.80	0.03	0.55	0.00		0.03	23.80	DCM(kmol/h)
TOTAL (kmol/h)	316.00	316.00	288.70	27.30	14.44	274.27	14.41		0.03	27.30	TOTAL (kmol/h)

Table 2. Mass and Energy Balances for the production of 7 200 kg/year of b carotene.

From the energetic point of view, the main drawback of the process is to achieve the separation of solvent and antisolvent without releasing too much pressure (because of the recompression costs). To avoid a solution with a high concentration of antisolvent inside the solution vessel, a purification of the solvent recirculation stream is needed. So for this step it is considered a flash cascade or a flash with a subsequent distillation tower where in the bottom, nearly pure liquid dichloromethane is recovered. But if just one flash is used, the concentration of antisolvent in the solvent recirculation stream is too high (close to 20%_w) to let the dissolution of β carotene in the feed stream. So, an extra purification step for the solution stream is required. Several alternatives were studied (a flash cascade to atmospheric pressure; a flash and a distillation tower at different pressure conditions), but the alternative chosen is a decompression to storage pressure with heating and consequent distillation. It is chosen because it presents a high purity on the recirculation streams without a high release of the pressure and subsequent recompression to storage pressure (as the flash cascade alternative). Besides, the working temperature of the boiler is not too high (it needs 10 bar saturated steam) and also the temperature of the condenser is not too low (cooling water can be used).

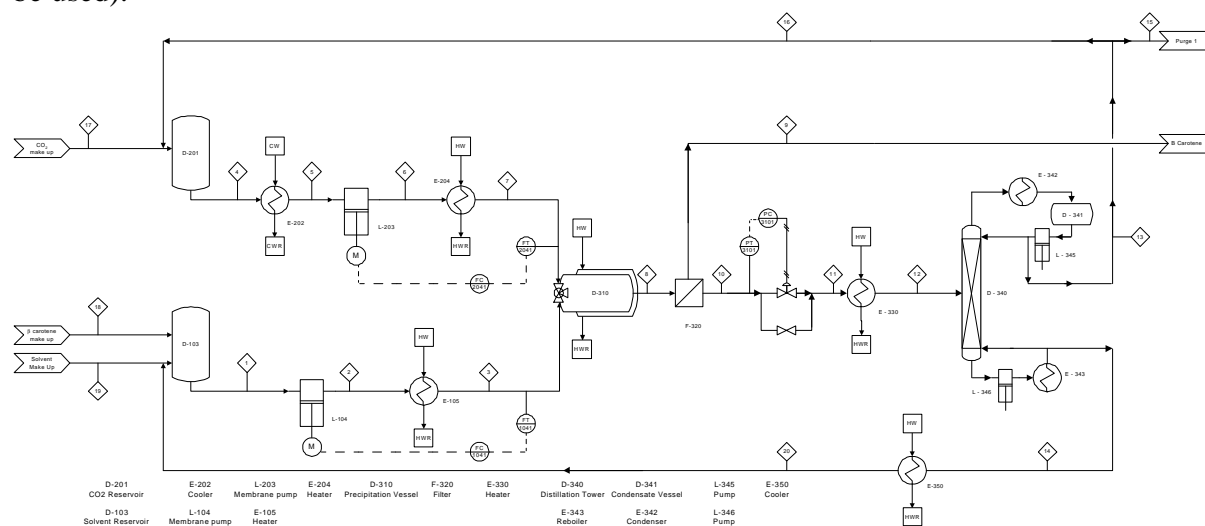


Figure 4. Flow diagram of for the production of 7200 kg/year of β carotene by PGA process.

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