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Study of process intensification Higee technology applied to the absorption.

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CONTENTS

SUMMARY

Higee technology or High-gravity Technology has been widely studied for the intensification of industrial processes such as absorption, desorption, distillation, extraction, among others, in order to find and improve the benefits compared to conventional methods. According to Documents about Best Available Techniques (BREFs), Higee technology is emerging as an alternative to the methods used so far.

This work studies Higee technology improvements applied to the absorption process compared to conventional packed columns. The absorption is mainly used to remove pollutant components of a gas stream to avoid leaks outside and/or to recover some component that the stream could contain for production purposes.

To achieve the goals of this project, a bibliographic search of Higee technology, studied by different authors, has been performed. In this research, the focus is on the modifications made in recent years in order to show the advantages of this technology compared to the conventional one.

After this research, the design of a conventional absorption column and Rotating Packed Bed have been studied in order to make a comparison among the design results and a comparison of these with those obtained experimentally by the different authors.

From these studies the main conclusion is that the use of Higee technology get a huge reduction in the volume of the device and it is also possible to use smaller amounts of absorbent liquid. However, Higee technology presents some disadvantages over conventional technology because it is necessary that the absorbent liquid be subjected to a centrifugal force, so this will require greater energy. All these changes make the design more complex and the device suffers a faster wear due to continuous movement and, consequently, it increases the economic cost.

Thus, based on studies performed so far, replacing the conventional absorption column for Higee technology is not recommendable. However, it should be further investigated because it has potential to get better results in comparison with conventional processes, either applied in the process studied in this work or applied to another types of processes.

RESUM

La tecnologia Higee o també anomenada tecnologia d'alta gravetat està sent àmpliament estudiada per la intensificació de processos industrials tals com l'absorció, la desorció, destil·lació, extracció entre d'altres, per tal de trobar i potenciar les millores que presenta respecte als mètodes convencionals.

Segons els documents que recullen les millors tècniques disponibles, anomenats *Best available techniques Reference document (BREFs)* Higee és una tecnologia emergent que representa una alternativa als mètodes utilitzats fins al moment.

En aquest treball s'han estudiat les millores que presenta la tecnologia Higee aplicada al procés d'absorció en comparació amb les columnes amb rebliment d'absorció convencionals. L'absorció s'utilitza principalment per eliminar possibles contaminants que hi pugui haver en un corrent gasós per tal d'evitar que aquests s'emetin a l'exterior i/o per recuperar algun compost necessari per a la producció.

Per aconseguir els objectius proposats s'ha realitzat una recerca bibliogràfica dels estudis de la tecnologia Higee desenvolupats per diferents autors i de les modificacions que s'han dut a terme durant els últims anys per tal de donar a conèixer els avantatges que presenta respecte a la tecnologia convencional.

Després d'aquesta cerca, s'ha realitzat un estudi del disseny d'una columna d'absorció convencional i del d'una columna rotatòria amb rebliment per tal de poder fer una comparació entre els resultats obtinguts i també una comparació d'aquests amb els obtinguts experimentalment pels diferents autors.

A partir d'aquests estudis s'ha arribat a la conclusió de que amb l'ús de la tecnologia Higee s'obté una enorme reducció del volum de l'aparell i també és possible utilitzar menor quantitat de líquid absorbent. No obstant això, s'ha comprovat que no presenta grans avantatges respecte a la tecnologia convencional degut a que s'obtenen uns rangs similars dels coeficients de transferència de matèria. A part d'això, és necessari que el líquid absorbent es sotmeti a una força centrífuga, de manera que aquest fet requerirà major energia. Totes aquestes variacions fan que el disseny de l'equip sigui més complex i pateixi un desgast més ràpid degut al continu moviment i com a conseqüència s'incrementa el cost econòmic.

Així doncs, a partir dels estudis realitzats fins al moment, la substitució de la columna d'absorció convencional per la tecnologia Higee no es veu afavorida. No obstant això, s'hauria de continuar investigant ja que presenta possibilitats d'obtenir resultats més òptims en comparació amb els processos convencionals, ja sigui aplicant-la en el procés estudiat en aquest treball o aplicant-la en un altre tipus de procés.

1. INTRODUCTION

Higee Technology or High-gravity Technology appeared in late 1970s but the first patent published appeared in 1956 from which allows the evolution of that technology adding different enhancements to obtain greater performances.

This technology is used for the intensification of industrial processes as adsorption, desorption, distillation, extraction and stripping among others.

Higee allows us to reduce the size and weight of the equipment, lowers capital costs, enhances the performance of the process, reduces energy consumption and minimizes environmental impact.

This technology mainly consists to replace the gravity fields to centrifugal fields. This method has been developed as alternative to conventional packing columns to enhance mass transfer between gas and liquid. These types of columns are called rotating packed bed (RPB) that it is an easy operation and it has a high efficiency.

Rotating packed bed was developed by Pilo Claes Wilhelm in 1956 (Claes Wilhelm P., 1956), but he is not recognized as the first inventor. In all the articles, the author who patented the Rotating Packed Bed was Colin Ramshaw in 1978 (Ramshaw, C., 1978).

The arrangement contains a rotor which has a packing that provides a large surface area and it is covered with a fixed housing. When this rotor runs on a fixed shaft, it provides a centrifugal field to the system.

The procedure consists on the inlet of the gas in the device by one side of the casing and goes inward through the packing due to the pressure difference until the upper part of the casing where there is the gas outlet. The liquid input by a liquid distributor situated at the center of the system. It goes outward to the casing due to a centrifugal and gravity field and emerges to the bottom of the casing where there is a liquid outlet.

When the liquid gets into the equipment it is uniformly distributed for all the system as small droplets and/or very thin film that provides a larger superficial area contact between the gas and the liquid than a conventional column because in this one, the liquid forms a thicker film and consequently there is less contact area.

When the liquid and gas are in contact, they can exchange mass between them. The mass transfer involves transfer of a solute from the gas to the liquid or vice versa.

Comparing the conventional column with Rotating Packed Bed there is an enhancement in the mass transfer coefficient. Therefore, with the application of Higee Technology we can have larger mass-transfer velocity between the gas and liquid.

From other side, Higee Technology is suggested as an Emerging Technique for Manufacturing of Organic Fine Chemicals of the Reference Documents about Best Available Techniques (BREF).

For these reasons, the study of this technology is very interesting because it can offer many advantages in comparison to the convectional technology.

2. OBJECTIVES

The objective of this project is to study Higee Technology to see the progress until nowadays and analyze possibilities of substituting conventional absorption columns. In order to achieve this objective, the following works will be carried out:

- A literature review of scientific articles and patents with the aim of comparing the information of different authors and contrast their points of view.
- Select mathematical models for designing conventional and Higee absorption and apply them to a common absorption problem
- Compare design results for the two technologies and discuss about possibilities of substituting conventional technology.

The bibliographic study about evolution of Higee technology is presented in the following chapter 3, the design of the equipment corresponding to the two technologies is presented in chapter 4, while comparison of results and discussion about possibilities is presented in chapter 5. Finally, in chapter 6, main conclusions of this work are summarized.

3. DESCRIPTION AND TECHNIQUE EVOLUTION

According to the articles that we can found, Higee Technology was patented in 1970 by Colin Ramshaw (Ramshaw, C., 1978), but according to the bibliography source, the first patent which the operation is described, was published by Pilo Claes Wilhelm in 1956 (Claes Wilhelm P., 1956).

3.1. DEVICE DESCRIPTION

In the aforementioned patent describes apparatus for the performance of an exchange of heat and/or soluble substances between two different media that were flowing of different specific gravity.

Until now the design of the apparatus for the exchange of substances and/or head is based on obtain a large contact surface per unit of volume between two medias, but Pilo Claes Wilhelm, in the aforementioned patent, proposes that it is also important to consider the amount of heat and/or solute transferred on the contact surface. Therefore, it proposes an apparatus that its capacity is not only based on the contact surface per unit of volume, but also it is important take into consideration the interface contact between the two fluids.

So, as Claes Wilhelm indicates *"may have better results if the operation is carried out with the greatest possible relative velocity between the medias, producing the greatest possible turbulence in the gas side and changes it direction and if in the liquid side is given a high speed so that it spends in thin layers on the support."*

According to this invention, the process involves contacting two medias, which one of this medias is heavier and the other is lighter, subjecting them to centrifugal acceleration. This centrifugal acceleration is between 15 and 200 times the acceleration due to gravity in agreement with aforementioned patent.

This happens in apparatus comprising a housing (1b) where inside of the latter there are a rotor which is forming with an annular element (3) with a large interfacial area, which is hooked into a shaft (4) and them rotate to provide the centrifugal acceleration. The heavier medium is distributed covering the walls of the annular element with a very thin layer. According to this patent, the thickness of this layer is between 0.1mm and 0.01mm or less, and it contacts with the high velocity flow medium lighter.

The process method is as follows: the gas enters in the inner space (1a) of the housing (1b) through a tube (1). This gas passes through the rotor (3) which is formed by an imperforate base (6) fixed to the rotation axis (4) and a cover (3a) which also it is not perforated. The base and the cover are held together by screws (12).

Inside the rotor (3) there is a packing (10) that can be formed by metal wire or grids, but also instead of packing, there can be a set of metallic plates (11) arranged in parallel and close together. There can be different types of plates and dispositions of the same, these will be mentioned further on.

The packing (3) allows to break the gas flow and it is distributed in thin layers throughout the packing. Therefore, the gas will pass thought the whole packing until arrived at the intermediate chamber (3b) who communicates with the gas outlet tube. (2).

The fluid enters through a pipe fixed (15) located on the top of the housing (1b). This one passes through the perforated plate (16) and it is distributed inside of the screen (17) and by the dished member (13). From here, the liquid is projected toward outside and distributed through the plates or packing (3), as mentioned previously, on thin films in such a way that gets a larger interfacial area that allows a greater mass transfer between the gas and liquid with a shorter time of contact. According to this patent, the contact time range is between 1s and 6s.

Hereafter there are two diagrammatically of the device:

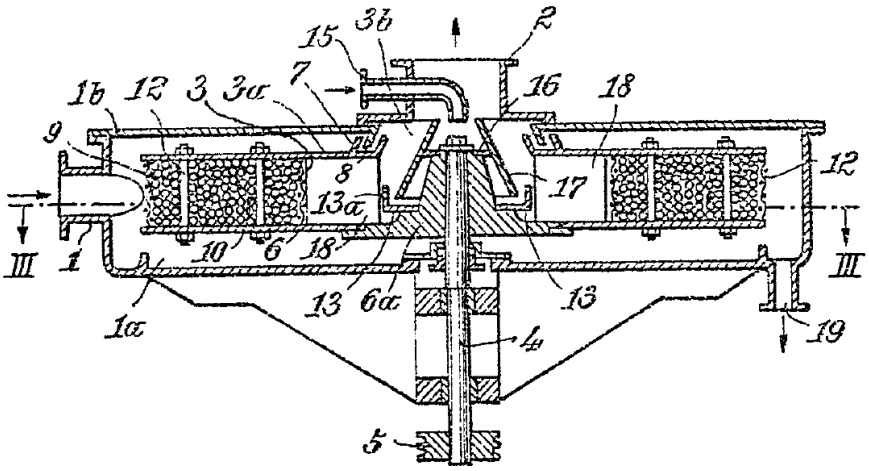

Figure 3.1.: Scheme of the vertical section of the device according to the Patent No. 757.149 (Claes Wilhelm P., 1956).

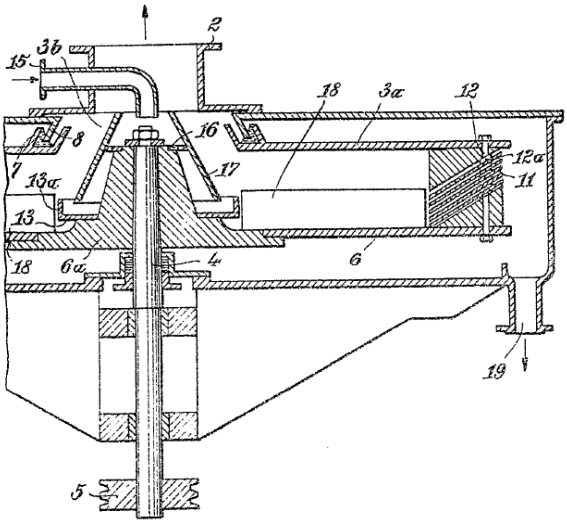

Figure 3.2.: Scheme of a section to a large scale of another type of device as a patent No. 757.149 (Claes Wilhelm P., 1956).

Pilo Claes Wilhelm also propose, perform the procedure at several stages by connecting more than one apparatus in series or in parallel to achieve removal different component, but it not entered in detail because, connecting several equal devices, is not the main idea of this work.

Later, in 1969 David B. Todd patented a gas-liquid contactor multistage (D.B.Tood, 1969). Probably he took the idea proposed by Pilo Claes to carrying out the procedure in several stages, but instead of connecting in series or in parallel more than one device, he decided to join the rotors in one shaft to form the apparatus and he called it multistage contact. This is explained in detail in section 3.2.

In 1978 Colin Ramshaw proposes a rotatory apparatus for a mass transfer. It comprises a permeable element, which it has an annular or disk form, so that it has a large interfacial area in such away it can achieve a greater increment of mass transfer (Rashaw C., 1978). This apparatus is applied to the company Imperial Chemical Industries Limited.

In my opinion, the apparatus proposed by Ramshaw is approximately the same that was proposed by the above-mentioned Claes-Wilhelm Pilo, but there is a hypothesis that an English company called Imperial Chemical Industries bought Pilo's idea in order to have a good reputation in the chemical sector.

Colin Ramshaw introduced several important concepts about the process, that it had not been mentioned in the patent described before. These concepts are explained below.

One of the most important concepts that Ramshaw introduced regarding to the mass transfer are mass transfer coefficients of the liquid and gas which are K_L and K_G respectively. According to the values of that constant, the process can be controlled by the liquid or gas phase.

The gas phase controls the process when the mass transfer is limited by the solute diffusion through the gas phase, and therefore, the liquid phase controls the process when the mass transfer is limited by the solute diffusion through the liquid phase.

In several studies, he saw that the values of mass transfer coefficient in the liquid phase and in the gas phase are greater than expected if the fluids are subjected, during the mass transfer, to a high acceleration at least 300 m/s^2, according to this patent. Hence, the use of Higee Technology encourages the mass transfer process.

According to Ramshaw, it can be two types of contacting between the two fluid phases, in co-current or in counter-current. He concludes that if the rotation speed of the permeable element is high, then it provides the counter-current flow, instead, if the first fluid flow is high then it provides the co-current flow. However, is preferable work with counter-current flows.

Also he introduces the residence time. The fluid residence time within the permeable element is in function of the permeability and radial dimensions of this one, the rotational speed and flow rates of fluids.

The permeable element and/or packing must be composed by an agglomeration of fibers that it can be formed for monofilament with diameter less than 150 microns (Ramshaw C., 1978). These have to roll slightly under tension around the shaft to forming the annular element that must have a minimum porosity between 90% and 93%, in order to achieve greater interfacial area and consequently greater mass transfer. The wetting of the permeable element depends on the fluid dynamics. To improve wetting can be applied in the packing, a wetting agent or add the wetting agent directly in one each of the fluids.

In addition to this, the material that formed the permeable element must have a specific feature. It must have mechanical resistance to withstand the tension generated in the material due to the rotation speeds at which the apparatus is subjected without breaking. This also has to be resistant to the possible reactions of fluids that will be in physical contact.

Another parameter that it is important to consider is the pressure drop. This concept has not been studied in detail in patents described above. The pressure drop in the rotor increases due to centrifugal forces and friction because of the gain of the kinetic energy (Kishore, P., 1990).

The experiments realized in 1996 (Hwai-Shen, L., 1996) revealed that, if the rotor does not contain liquid, the relation between the pressure drop and the rotation speed is linear to different gas flow rates. It also was observed that the pressure drop with high gas flow rates is affected lightly by the liquid flow. Therefore, as mentioned earlier, in the patent No. 0.002.568A, at high rotation speeds, the gas flow affects the pressure drop.

From these devices, many studies have been carried out in order to improve device performance, mainly to increase the mass transfer ratio. To accomplish this, some modifications on the apparatus have been done, on the packed bed and on the type of fluid contact, which are explained below.

3.2. DEVICE EVOLUTION

Contactor multistage

As been mentioned before, in 1969 David B. Todd patented a gas-liquid contactor multistage (D. B. Tood, 1969).

The device to carry out this procedure consists of a tank (10) which is divided in three areas (11, 12, 13) which each one contains a rotor (33, 34, 35), where the liquid and the gas will get in contact. The rotors are attached to the shaft (14) through support collars (55, 56, 57) located at the bottom of the rotor (33, 34, 35). Each of these areas is separated from the other by a partition that has a certain downward slope inward (18, 19, 20). These partitions are removable because they are joined to flanges (21, 22, 23) which are fixed to the wall of the tank.

Each rotor (33, 34, 35) contains an upper wall (39, 40, 41) and an inferior wall (36, 37, 38). The upper walls contain an opening central (42, 43, 44) which are connected the vapor passage (27, 28, 29). These steps around the shaft (14). They have an annular cross-section and they are bounded by tubular baffles (30, 31, 32) with a curved outwardly-extending lips to facilitate fluid flow and maintain the liquid and the gas separated.

Within each rotor there are a plurality of cylindrical elements (45), with many holes (46), arranged in parallel to the shaft (14). Right in the center of the rotor there is a space to gather the steam (47, 48, 49) that it connected with an opening central (42, 43, 44) abovementioned.

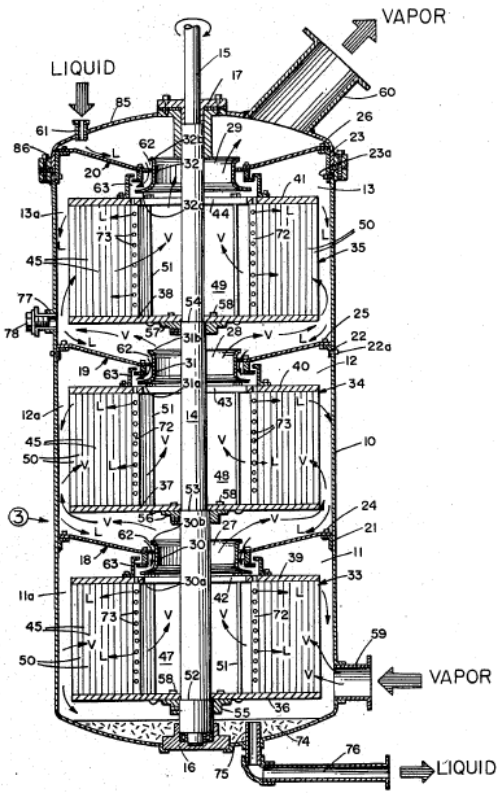

Figure 3.3.: Schematic of multiple gas-liquid contact apparatus according to Patent No. 3.486.743 (D.B.Tood, 1969).

The procedure consist that the gas is delivered to the first zone (11) by a pipe (59), which is located in the lateral bottom of the tank. The gas passes through the first rotor (33) from the periphery to toward the center of the tank and it leaves of the rotor by vapor passage (27). The gas is following the same path by the second and third rotor (34, 35), up to coming to the upper space of the third partition (20) which is connected to the gas outlet (60).

In the third space partition apart from having the outlet gas, aforementioned, also there is the liquid inlet (61). When the liquid enters moves through the third partition (20) and through the passage (62) it enters in the third rotor (35) where there are the cylindrical elements (45). The liquid moves from the center of the rotor to the periphery of the tank. It is following the same path by the second and first rotor (34, 33), up to coming to the bottom of the tank (20) where

there is the outlet liquid (70). Also it is necessary to mention that, in the apparatus aforementioned, can has a second inlet liquid (77) which it is located above to the second rotor.

Using this method, in which there are three rotors connected in series, it is possible to maintain continuously in counter-current contact the liquid and the vapor flow.

Cross-flow rotating bed

In 1997 Fen Guo et al. proposed to modify the rotor in order to change the type of contact between the fluids. Until now, it was observed that if the fluids are in counter-current contact was achieved a greater mass transfer.

They suggested contacting the two fluids in a crossflow rotating packing bed. To achieve this, it was necessary to modify the entries of the gas and the liquid. In this case, the liquid enters through the side wall of the casing and it is distributed by the packing through the liquid distributor moving in the direction perpendicular to the axis. Instead, the gas enters through the top of the casing and it is distributed by the packing in the direction parallel to the axis so that contact between the liquid and the gas was cross-flowing.

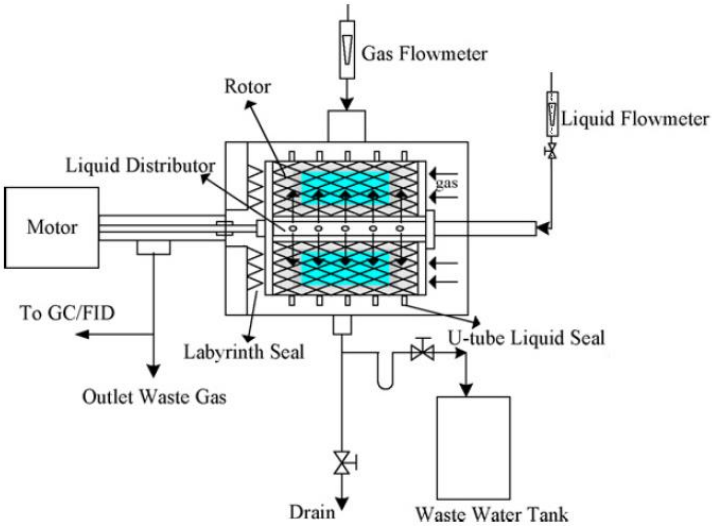

Figure 3.4.: schematic diagram of Cross-flow Rotating Packing Bed.

Using the results, it was a great improvement with regarding to the counter-current rotating packed bed and the conventional packing. This type of multiphasic contact allows reducing the size of the rotating packed bed in comparison with other types of contact of fluids. It also was

observed that there is a significant increased mass transfer coefficient by increasing the liquid flow.

After this invention other authors have been published articles where they modified the type of the packing to enhanced the cross-flow contact between the fluids. This is explained in the section 3.3.

Multi-rotors zig-zag rotating beds

In 2008 Ji et al. patented a Multi-rotors zig-zag rotating beds. (Jianbing, J., 2008) The system procedure is similar to the device patented by David B. Todd (D. B. Todd, 1969), but it changes the type of the packing. The packing proposed allows the contact between the gas and liquid is in zig-zag. Rotating zig-zag bed differs from the others packings since it is the only one that contains a combination of a rotating element with a stationary element. This device allows use an easily intermediate liquid inlet so that it can perform a continuous process.

The objective of using multi-rotors contact is likely as the patent aforementioned, but here the main objective is increasing the mass transfer capacity between the gas and liquid to obtain a further throughput of the apparatus.

The device described in this patent is formed by a casing (8) which in the center of thereof there is a rotatory shaft (10) that it goes from the upper to the bottom of the casing. At the top of the casing (8) there are, as in the most of the devices described in the bibliography founded, the liquid inlet (7) and the gas outlet (5) and at the bottom of the casing (8) a gas inlet and liquid outlet. In the shaft (10) there are fixed three rotors mounted in series.

As shown in Figure 3.7, each rotor is formed by two discs. The upper disc is fixed (4) and the lower disc is mobile (1). Between the two discs there are a set of channels arranged in zig-zag shape (13).

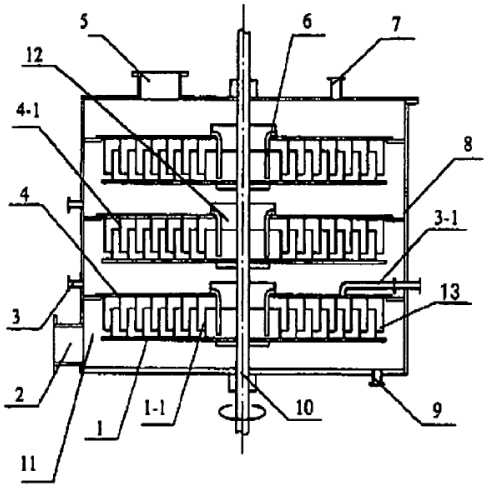

Figure 3.5.: schematic diagram of multi-rotors zig-zag according to the Patent No. 7.344.126 B2 (D. B. Todd, 1969).

This method consists in the entrance of the gas in the casing (8) through the gas inlet passage (2). After, the gas passes through the first rotor starting from the periphery of the casing to inwardly passing through zig-zag channels (13) in spiral way under a pressure difference and leaves the rotor through the hole located at the center of the upper stationary disk. The gas performs the same path through the second and third rotor until arrive at the upper part of the case where there is a gas outlet orifice (5). There will also be, as mentioned before, the liquid inlet (7). When the liquid enters to the casing, it moves above to stationary plate of the three rotors. Then it enters to the rotor through tube (6) located in the center hole of the upper disk. From there, the liquid will be projected into small droplets to the channel zig-zag. The centrifugal force makes that the liquid flow from the center of thee rotor to the periphery of the case. The liquid performs the same path through the second and first rotor until arrive at the bottom of the case where there is a liquid outlet orifice (9). The liquid, a part of the upper inlet (5), it can also have other inlet pipes (3) located in the wall side of the casing.

The author of this patent proposed different shapes of the packing rotor, these are mentioned in the section 3.3.

In a rotating packed bed are distinguishable three main areas where the mass transfer takes place, the end effect zone corresponding to the inner edge part of the rotor, the bulk zone

corresponding to the rest of the packing of the rotor and the cavity zone, which corresponds to the area between the rotor and the casing.

It was observed that the end zone allowed to obtain a greater mass transfer than in the bulk zone such as shown in Figure 3.10, these were the results obtained (Yong, L., Guang-Wen, C., 2012).

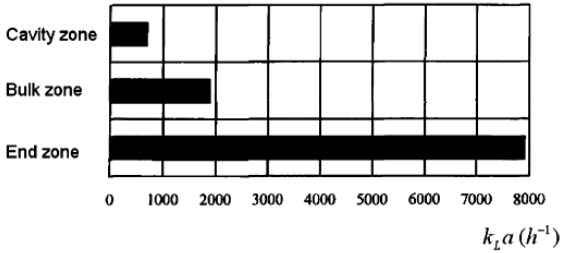

Figure 3.6.: Graphic of mass transfer rate in each zone in the rotating packed bed.

Therefore, it was found that in the end zone the mass transfer has more important than in the bulk zone. This is due to various factors, according to Luo Y.: *"occurrence of the most violent collisions between incoming liquid jets and rotational packing with maximum relative velocity, the strongest gas and liquid interactions, because of the maximum gas flux rate across a minimum cross-sectional area of the end zone and the most rapid resupply of fresh liquid from the liquid distributors"* (Yong, L., 2012).

For this reason, Guang-Wen and Yong, L, C. et al. proposes Multi-inlet Rotating Packed. (Guang-Wen, C., Yong, L., 2014).

Multi-inlet Rotating Packed

Multi-inlet Rotating Packed comprises a rotor formed by three rotating rings which have different radius. The liquid is introduced through the pipes Q11, Q12, Q13 in each rings aforementioned. Every ring contains four liquid distributors arranged concentrically. Through these holes the liquid is sprayed outward through the rotor, due to the centrifugal force, to reach the bottom of the area between the rotor and the casing, where is the liquid outlet (1).

The gas enters through the orifice (2) located tangentially and in counter-current respect to the liquid inlet. This is moved into the rotor due to the pressure difference, until it arrive at the center, where is the gas outlet (5).

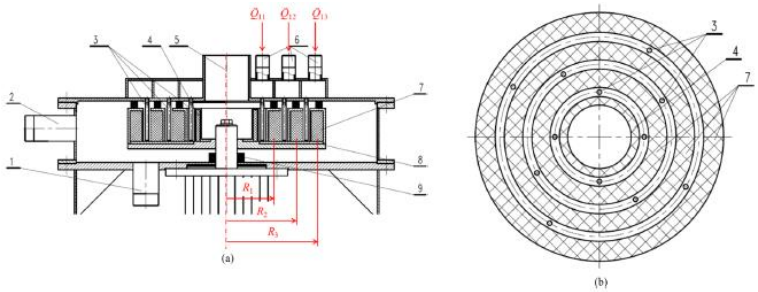

Figure 3.7.: Schematic diagram of Multi-Inlet Rotating Packed Bed.

Multiples inlet liquid, distributed by rotating rings, allow provide a continuously fresh liquid into the rotor rings.

3.3. PACKING EVOLUTION

Rotating Packing Bed

As it was mentioned before, in the description of the first apparatus patented, (Claes Wilhelm P., 1956) it is possible to use several types of plates and dispositions of themselves. The plates can be available in parallel to the shaft of rotation, as shown in figure 3.3, they can be inclined in relation to the shaft, from 0º and 90º, as shown in the figure 3.4 and also they can be disposed inclined in relation to the vertical plane as shown in figure 3.4, plates 11f.

The plates can have different shapes. They can be flat (11a, 11c, 11d, 11f) or also they can have zig-zag shape (11b, 11e), the disposition of which is a narrow passage between them. To make the passage between the plates of the same size can be made thicker plates to the periphery (11g). It may be advantageous that the plates are curved on the inside in the direction of rotation (11h) since it makes the distribution of the liquid faster and they can also be arranged at right angles along the periphery of the rotor (11d) plans.

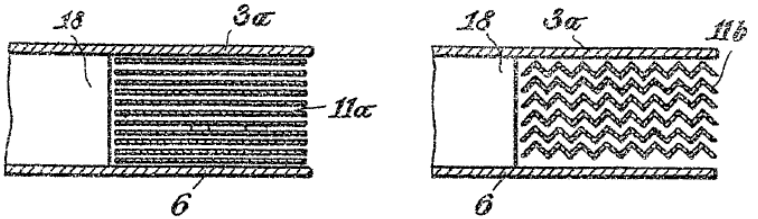

Figure 3.8.: parallel disposition of the plates as the patent No. 757.149 (Claes Wilhelm P., 1956).

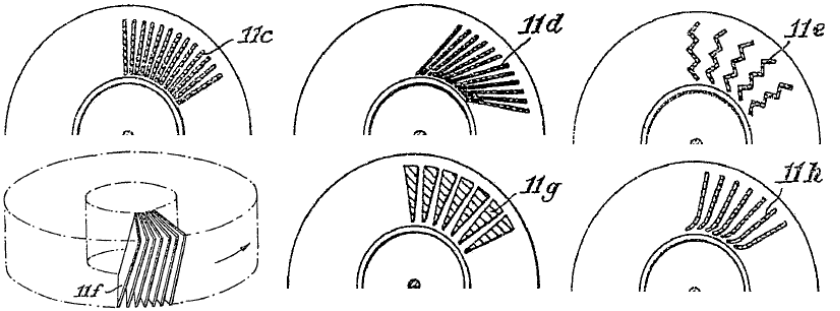

Figure 3.9.: Disposition of the pates inclined to the shaft according to the patent No. 757.149 (Claes Wilhelm P., 1956).

Split Packing

In the article published by A. Chandra is about Higee technology applied on a rotor with a Split packing (A. Chandra, 2005) to intensify mass transfer and to improve slip velocity, since the slip velocity between the gas and liquid phase is approximately the same as conventional columns.

To achieve these objectives, it proposes divide the packing into equidistant annular rings and alternate rotation in counter-direction. As the packing has spaces between the rings, this allows that the liquid is sprayed from one ring to another, and therefore to achieve an improvement of the transfer on the liquid side phase.

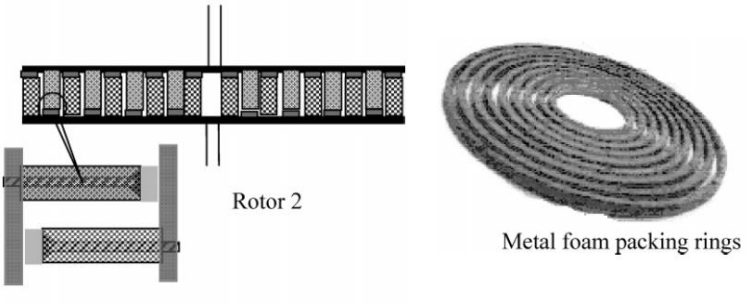

Figure 3.10.: Schematic diagram of a Split packing.

It was observed that, with good equipment design and working in countercurrent, mean split velocity is achieved, according to the article the split velocity is between 10 and 25 m/s. That, theoretically, allows enhancing the mass transfer in the gas side, so this leads to achieve high performance.

With this study, it was observed that the split packed could allow the improvement of the mass transfer ratio, but according to the article, more experimental confirmations are necessary. K. Jagadeswara performed more experiments with this type of packing. (K. Jagadeswara, 2006).

K. Jagadeswara realized a study with the objective to evaluating the effectiveness of a split packing for enhance the mass transfer.

The results that they obtained were compared with bibliographical information and with the data of the conventional columns. They concluded that the results obtained do not shown a great improvement, therefore, this type of rotor does not compensate, because there is no improvement in the rate of mass transfer large enough to compensate the complex device, because it probably has to have two motors, one for each rotating annular rings and consequently it implies a greater energy consumption.

Multi-rotors zig zag bed packing

The author of this patent proposed different shapes for the packing that are shown in the Figure 3.8 the packing of the rotor can have different shapes: the embodiment can contain a set of holes (a), also instead of holes it can be opened in a jalousie. This can be open in the clockwise or counterclockwise (b) and finally, it maybe has a mixture of the two packings mentioned, it can be opened in a jalousie and under these it can have a line of holes.

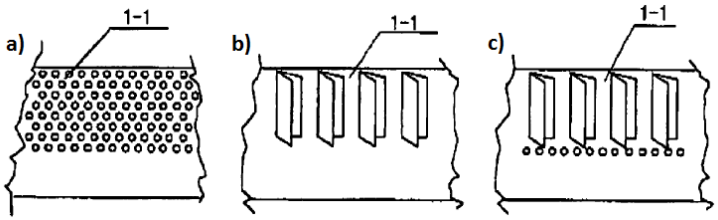

Figure 3.11.: different packing shape of the zig-zag packing.

G. Q. Wang et al. realized a series of experiments, using the rotating zig-zag bed. They concluded that the efficiency of the mass transfer obtained with this apparatus is approximately equal to that of the others apparatus aforementioned, but in the case where a multi-rotor is used the mass transfer capacity increases significantly.

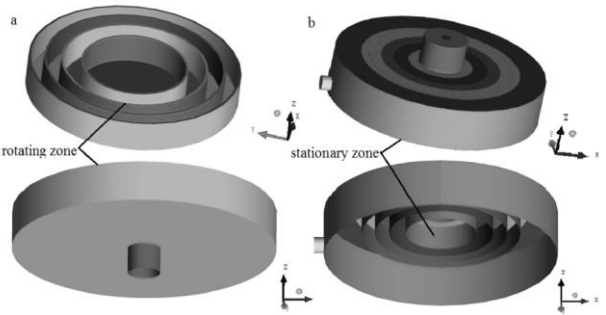

Figure 3.12.: schematic diagram of rotating (a) and stationary (b) zig-zag rotor (Sun Y., 2014).

Cross-flow rotating bed

As noted that the cross-flow not showed improvements in the mass transfer coefficient, a number of modifications to the original packing crossflow were made for obtain greater mass transfer coefficients in comparison with the rotating zig-zag bed.

In 2009 Guang W. et al. decided to carry out the study of the cross-flow performance in a Higee no packing. To do this, they mixed the idea of cross-flow with the zig-zag rotating packed bed proposed by Ji Jianbing (Jianbing, J., 2008).

In this case, the rotor consists of perforated rotating disk by a set of large holes through which passes the gas flow. This rotating disk also has a set of fix rings with small holes at the top through which passes the fluid flow. In this way there is continuously axially contact between the liquid and the gas, since the gas moves in parallel in the direction of the axis of rotation and the liquid moves perpendicular to the axis of rotation.

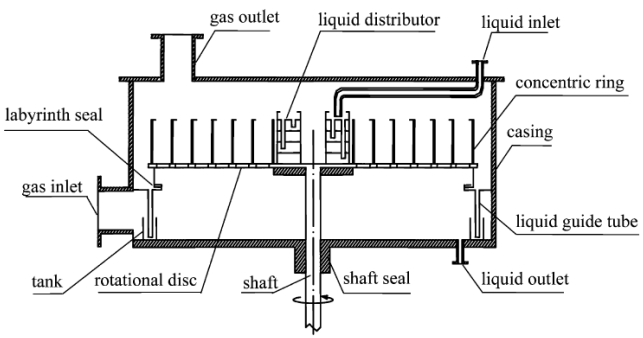

Figure 3.13.: schematic diagram of Cross-flow Rotating Bed with rotational baffles.

With the obtained results it is observed that the mass transfer coefficient has an order of magnitude less than the counter-current flow and that the cross-flow rotating packed bed.

A few years later, in 2014, the same authors of the article mentioned above, decided to modify the structure of the rotor. They wrote an article that they divided in two parts. In the first part, the rotor consists of a rotating disc and a stationary disk. Between them there are a set of rotational baffles that are divided into three zones, on one end there are no holes, in the central part there are small holes for the passage of the liquid fluid and the other end contains large holes for the passage of the gas fluid. This allows the gas flow moves through the rotor by the zig-zag channels and the liquid moves in a parallel flow path to the rotating disks, so there is a repeated contact between the liquid and the gas, as seen in Figure 3.14.

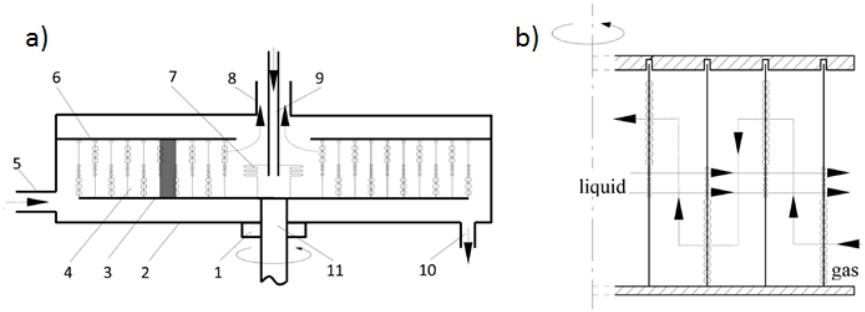

Figure 3.14.: Schematic diagram of Cross-flow Rotating Bed with zig-zag gas flow.

The results obtained in this type of rotor show that the Cross-flow Rotating Bed has a poorer mass transfer in comparison to Rotating zig-zag Bed. Although cross-flow presents advantages in pressure drop and shaft power, but it is not enough. They concluded that the mass transfer performance can be affected by three factors: liquid maldistribution, the flow model and the seal.

In the second part of this article, the rotor consists in a stationary disk, that it is an upper disk and a rotating disk, that it is a lower disk. The latter contains a series of concentric baffles that extend into the seal located at the stationary disc so that form a labyrinth seal between them.

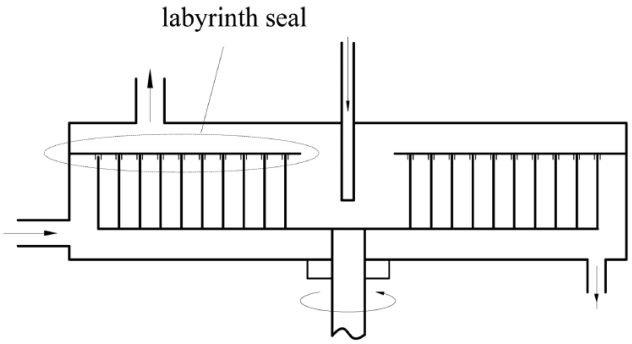

Figure 3.15.: Schematic diagram of Cross-flow rotor with gland seal.

This type of rotor allows contact between the gas and the liquid in cross-flow. To this, the article proposes three different types of baffles, which they can see in the Figure 3.16.

The first one is the same mentioned in the above patent. In this the liquid fluid path is perpendicular to the axis, while for the other two baffles liquid follows a path in a zig-zag. This is achieved by decreasing the rows of small holes, through which pass the liquid. As can be seen the rotor 1 has 14 holes, whereas the rotor 2 and 3 have 7 and 3 holes respectively.

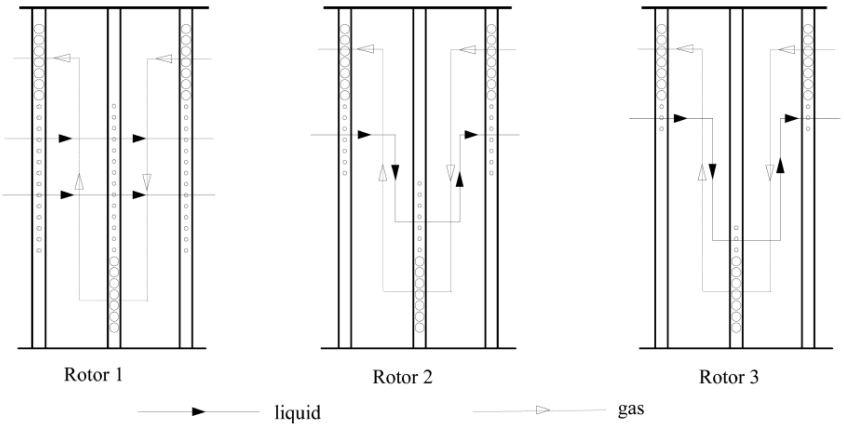

Figure 3.16.: Schematic diagram of liquid and gas contact in different Cross-flow rotors.

With different tests on these types of rotors, they concluded that the cross-flow contact provides less efficiency with greater energy consumption and an increase of the pressure drop. Thus, after several attempts it was clearly observed that the contact between the fluids zig-zag contact is better than in cross-flow.

4. MATHEMATICAL MODEL AND DESIGN

4.1. MATHEMATICAL MODEL AND DESIGN FOR CONVENTIONAL PACKED COLUMNS

Absorption is a unit operation of mass transfer widely used in the chemical and petroleum industry. The main objective of the absorption is to separate one or more minority components called solute, from a gas mixture with the help of a liquid stream. The liquid and gas must be immiscible.

The absorption can be chemical, if there is a chemical reaction between the solute and the absorbent or can be physical if there is no chemical reaction between them. The absorption is reversible; its invers process is called desorption.

The mass transfer between the gas and liquid takes place through the contact surface due to concentration gradients as the driving force. Mass transfer depends on the interfacial area providing the liquid; therefore, it is important to note that the apparatus used encourage the gas-liquid contact and consequently providing maximum transfer area.

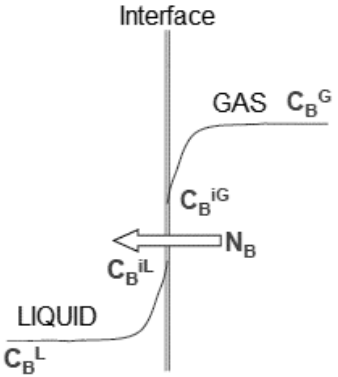

Figure 4.1.: Representation of the mass transfer between the gas and the liquid.

Mainly the absorption is used to remove possible pollutant components of the gas stream to avoid leaks outside or to recover some components that the stream could contain for production purposes.

The most commonly device used are plate columns and packed columns where the liquid falls down due to gravity force from the top of the column to bottom and the gas moves from the bottom to the top of the column due to pressure difference.

Another device that it is used in the USA but it is not available in Europe, according to Dr. Mark Roelands, is Higee Technology. This technology applied to absorption columns, as it has been mentioned before, uses centrifugal force to disperse uniformly the fluid and the pressure difference for passing the gas through the column.

In this section are developed the mathematical models for the design of the conventional column and for the design of the Rotating Packed Bed, in order to obtain comparable results. For this reason, a conventional packed column has been chosen, because it is more like to Rotating Packed Bed and the results obtained are comparable.

The process that has been chosen is the removal of carbon dioxide with water, because are provided all the necessary data and thus the results will be more real and clear for then compare the results more easily. The mathematic models are showed below.

4.1.1. Mathematical model

An absorption column design depends on the relationship of the flow rate and concentration of the inlet and outlet gas and liquid streams, the speed of absorption and mass transfer coefficients.

Then is shown, schematically, a packed column with the currents:

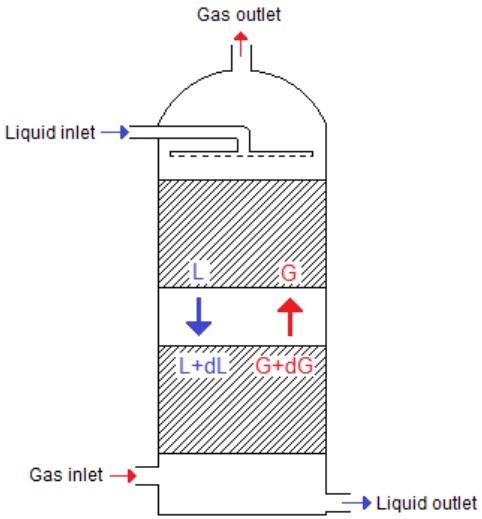

Figure 4.2.: scheme of a conventional column.

The tower consists on a liquid inlet and gas outlet, located in the upper part and a liquid outlet and a gas inlet located at the bottom of the column. Inside it contains a porous packed which provides a large contact area between the liquid and gas. The liquid will enter into the column, it can be pure or with a low concentration of solute, and it will be enriched as it descent through the packed due to the contact with the gas which is rich in solute. The gas goes up through the packing. Therefore, the mass transfer between the liquid and gas phase take place within packing.

To obtain a better contact between the liquid and gas fluid it is necessary that the liquid flows as a thin film on the surface walls of the packing to have a maximum contact surface area and therefore a greater mass transfer.

For this reason, it is very important packing characteristics in the design of the column since the mass transfer depends of it. In this work random packings will be studied, their characteristics are found hereinafter.

Mass balance in the absorption column

The global mass transfer in overall column is defined as:

$$G_1 + L_2 = G_2 + L_1 \tag{1}$$

where, G_1 and G_2 are inlet and outlet gas flow rates respectively and L_2 and L_1 are inlet and outlet liquid flow rates respectively.

Therefore, the solute mass balance is:

$$G_1 \cdot y_1 + L_2 \cdot x_2 = G_2 \cdot y_2 + L_1 \cdot x_1 \tag{2}$$

where, y_1 and y_2 are inlet and outlet gas compositions respectively and x_2 and x_1 are the inlet and outlet liquid compositions respectively.

If the solute mass balance is performed in volume differential and it is assumed that liquid and gas flow rates are constant, reordering the equation, the called operating line is obtained:

$$G_1 \cdot y_1 + L \cdot x = G \cdot y + L_1 \cdot x_1 \tag{3}$$

$$y = y_1 + \frac{L}{G} \cdot (x - x_1) \tag{4}$$

where, x and y represent the overall liquid and gas compositions respectively in contact with each other.

To take place the absorption, the operating line must be above the equilibrium curve, as this provides a positive driving force, in the case shown below, the desorption would take place.

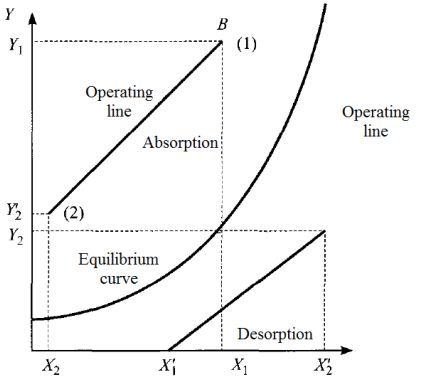

Figure 4.3.: graphical representation of the operating line and the equilibrium curve.

As aforementioned, in the interface between the liquid and the gas, there are a thermodynamic equilibrium. The equilibrium equation can be obtained numerically in two different ways, depending on the components:

When the gas that is in equilibrium with the liquid is ideal, hence, it follows the ideal gas law, Raoult's law is applied:

$$P_B = P_{v_B} \cdot x \tag{5}$$

For non-ideal liquid solutions, Henry's law is applied:

$$\frac{P_B}{P} = \frac{H}{P} \cdot \rho_M \cdot x \tag{6}$$

It is very important the value of the slope of the operating line, that is, the L/G, because when the operating line up to the equilibrium curve, and therefore, x_1 and y_1 are in equilibrium, the column needs infinite height. When this occurs, the L/G ratio is called $(L/G)_{min}$, because if working with a L/G lower than $(L/G)_{min}$, the driving force, that it is (y-y*) would be very small, and absorption would not be profitable. Therefore, liquid flow should be approximately 1,5 times L_{min}. This value is calculated from the expression obtained by rearranging the equation 4:

$$L_{min} = G\frac{(y_2 - y_1)}{(x_2 - x_1{}^*)} \tag{7}$$

During the process, there is a variation in the flow of gas and liquid due to absorption. These variations make the operating line curved，In the event that the system is diluted, that is to say, the flow rates remain approximately constant, these effects can be neglected.

To test whether the system is diluted, it will be necessary to verify that the following equations are fulfilled:

$$G' = G(1-y) = cte \tag{8}$$

$$L' = L(1-x) = cte \tag{9}$$

The mass balance of the solute in molar ratios will be:

$$G'\left(\frac{y_1}{y-y_1} - \frac{y_2}{y-y_2}\right) = L'\left(\frac{x_1}{x-x_1} - \frac{x_2}{x-x_2}\right) \tag{10}$$

Absorption rate and mass transfer coefficients

The absorption rate is expressed from a coefficient by the driving force. In this case, the coefficient used is the individual or global mass transfer coefficient in the gas or liquid phase, and the driving force is the difference in concentration of solute gas phase or liquid phase. Thus, they have different equations for calculating the absorption velocity, for calculation this from the individual mass transfer coefficients the following expression is used:

$$N_B\left[\frac{mol}{m^2 \cdot s}\right] = k_G \cdot P \cdot (y - y^*) = k_L \cdot \rho_M \cdot (x^* - x) \tag{11}$$

where, k_G and k_L are individual coefficients of mass transfer of the gas phase and the liquid phase respectively, x and y, are the molar fraction from the liquid and gas flow rates respectively and x* and y* are the molar fraction from the liquid in equilibrium with the gas or vice versa respectively.

The values of these coefficients can be obtained from the following expressions:

Gas phase individual coefficient:

$$\frac{k_y \cdot (Sc_G)^{2/3}}{G/S} = 1{,}195 \cdot \left[\frac{d_s \cdot {}^{G_m}/_{S}}{\mu_G \cdot (1-\epsilon_L)}\right]^{-0.36} \quad where \quad Sc_G = \frac{\mu_G}{\rho_G \cdot D_{BC}} \tag{12}$$

where, $k_y = k_G \cdot P$ and $\varepsilon_L = \varepsilon - \varphi_{L,W}$

Liquid phase individual coefficient:

$$\frac{k_x \cdot d_s}{D_{BA} \cdot \rho_M} = 25.1 \cdot \left[\frac{d_s \cdot {L_m}/{S}}{\mu_L}\right]^{0.45} \cdot (Sc_L)^{0.5} \quad where \quad Sc_L = \frac{\mu_L}{\rho_L \cdot D_{BA}} \tag{13}$$

where, $k_x = k_L \cdot \rho_M$

It is important to remember that L_m and G_m are mass flow rate of the liquid and gas respectively and those units are, kg/s.

From individual coefficients values and the slope of the equilibrium curve, it is possible to obtain the global mass transfer coefficients:

Gas phase global coefficient:

$$\frac{1}{K_G} = \frac{1}{k_G} + \frac{m}{k_L} \tag{14}$$

Liquid phase global coefficient:

$$\frac{1}{K_L} = \frac{1}{k_L} + \frac{1}{m \cdot k_G} \tag{15}$$

where, $1/k_G$ and $1/k_L$ are the mass transfer resistance in the gas and liquid phase respectively.

It may be the case that the value of k_G is much smaller than the value of k_L, then the value of m/k_L will be negligible compared to the value of $1/k_G$. Therefore, in this case, it is said that controls the gas phase, that is to say, the global mass transfer coefficient in the gas phase (K_G) is approximately equal to the individual coefficient of mass transfer in the gas phase (k_G).

It can also occur that the value of k_L is much smaller than the value k_G, then the ratio $1/k_L$ will be much higher than the value of $1/(m \cdot k_G)$. Therefore, the latter can be neglected in this case, it can be said that controls the liquid phase, that is to say, the global mass transfer coefficient in the liquid phase (K_L) is approximately equal to the individual coefficient of mass transfer in the liquid phase (k_L).

Design equations

To determine the volume of the column it is necessary consider at first, the absorption flow rate since the integral of this expression for the differential concentration, the column volume is obtained:

$$d(Gy) = k_G a \cdot P \cdot (y - y^*) \cdot S \cdot dh \tag{16}$$

$$\int_{y_2}^{y_1} \frac{dy}{(y - y^*)} = \frac{k_G a \cdot P}{G} \cdot S \cdot \int_{h_1}^{h_2} dh$$

$$\int_{y_2}^{y_1} \frac{dy}{(y - y^*)} = \frac{k_G a \cdot P}{G} \cdot S \cdot h$$

$$V = \frac{G}{K_G a} \cdot \int_{y_2}^{y_1} \frac{dy}{(y - y^*)} \tag{17}$$

where the term $\frac{G}{k_G a}$ is called volume of transfer unit of the gas phase (VUT), it has units of m^3 and the term $\int_{y_2}^{y_1} \frac{dy}{(y-y^*)}$ that it is called number of transfer unit (NUT), it is dimensionless number.

If the equilibrium and operation equations are lineal the integral can be approximated:

$$NUT_G = \int_{y_2}^{y_1} \frac{dy}{(y - y^*)} = \frac{y_1 - y_2}{\frac{(y_1 - y_1^*) - (y_2 - y_2^*)}{ln\left(\frac{y_1 - y_1^*}{y_2 - y_2^*}\right)}} \tag{18}$$

In the case of liquid, the design equation is:

$$V = \frac{L}{K_L a} \cdot \int_{x_2}^{x_1} \frac{dx}{(x^* - x)} \tag{19}$$

where the term $\frac{G}{k_L a}$ is called volume of transfer unit of the liquid phase (VUT), it has units of m^3 and the term $\int_{y_2}^{y_1} \frac{dy}{(y-y^*)}$ that it is called number of transfer unit (NUT), it is dimensionless number.

In the case of the liquid, if the operation equilibrium and operation are lineal, the integral can be approximated:

$$NUT_L = \int_{x_2}^{x} \frac{dx}{(x^* - x)} = \frac{x_1 - x_2}{\frac{(x_1^* - x_1) - (x_2^* - x_2)}{ln\left(\frac{x_1^* - x_1}{x_2^* - x_2}\right)}} \tag{20}$$

Normally instead of calculating the VUT is calculated the HUT (height of transfer unit), but in order to compare conventional column with the RPB it is convenient to calculate the volume.

It has been observed in the set of equations shown until now, the design of the column is also important the packed characteristics. Then, the packed types and characteristics are displayed hereafter.

The more common packing in absorption columns

As been mentioned before, it is important to consider the type of packing used since the optimum packing depends on operating conditions and on the desired concentrations.

It is possible to distinguish two main packing types: those who are randomly placed, these have small size, between 6 and 75 mm, those 25 mm or less are used mainly in laboratory scale and the other types that are ordered or structured, these have a size between 50 and 200 mm.

One of the important features that should have the packings are that they have to be inert because they cannot intervene in the absorption process. The most commonly materials used are ceramics, metal and plastic.

The following figure shows some examples of the most common packings that are used:

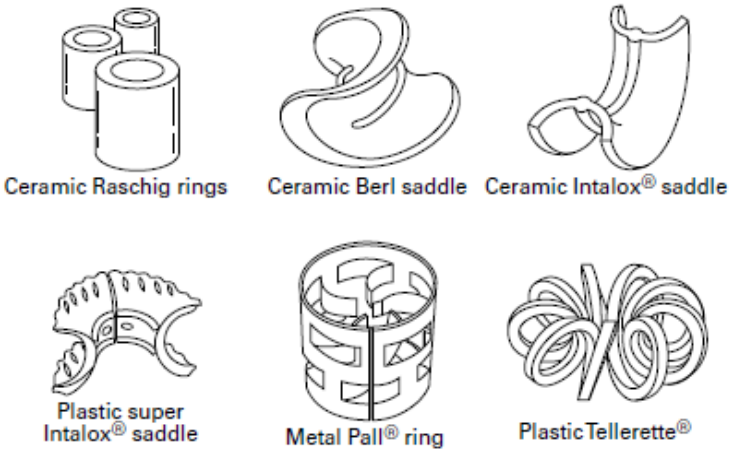

Figure 4.4.: Commonly used packings.

The characteristics of different packings types for different size are shown in the table A.1 (Appendix 1).

As mentioned above it is very important the contact area between liquid and gas, this is characteristic of the packing and is called specific area of the packing. As the liquid is not

distributed perfectly throughout the packing, it is important to consider in the column design, the packing wetting area. This depends on the mass flow of liquid and gas, the section of the column, the gas density and the type of packing. The expression to calculate the specific wetting area is:

In the case of aqueous liquids:

$$a_{AW} = m \cdot \left(\frac{808 \cdot {G_m}/{S}}{{\rho_G}^{0.5}} \right)^n \cdot \left(\frac{L_m}{S} \right)^p \tag{21}$$

For the case of non-aqueous liquids:

$$a_{AW} = a_{AW} \cdot \frac{\varphi_{L_o}}{\varphi_{L_oW}} \tag{22}$$

Consonant values m, n and p are shown in Table A.2 (appendix 2) and depends on the liquid flow divided by the section, L'.

The wetted interfacial area should be smaller than the dry area of the packing.

Another parameter that is important to know is fluid retention. This can be calculated from the following expression:

$$\varphi_{Lt} = \varphi_{Ls} + \varphi_{Lo} \tag{23}$$

The parameters to calculate the fluid retention are shown in table A.3 (Appendix 3).

Pressure drop

The pressure drop allows the pass of the gas through the column and consequently increases the capacity of the column and the mass transfer velocity. The pressure drop per unit of length of the packing is due to fluid friction. The pressure drop in the wet packing is greater than in the dry packing due to the liquid in the column reduce the space where the gas flows. Thus, it is important to considerer the packing porosity. Therefore, the pressure drop increases by increasing the gas flow. For this reason, to calculate the pressure drop within the column is necessary to consider that the column is wet. Then, for calculating the pressure drop the following graphic is used:

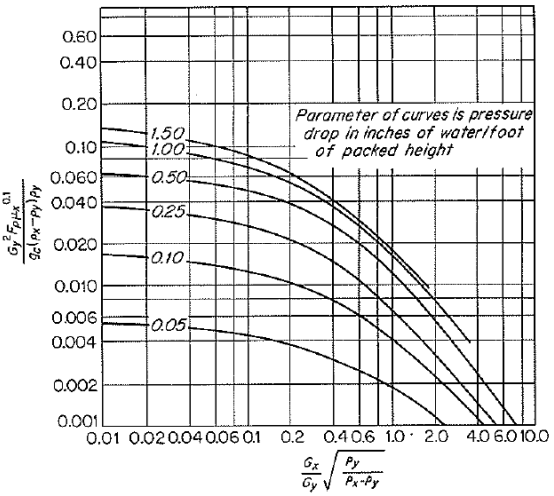

Figure 4.5.: Flooding and pressure drop in packed columns.

However, the pressure drop per unit of length considering that the packing is dry can be calculated with the following expression:

$$\frac{\Delta P}{Z} = C_D \cdot \frac{\left(G_m/S\right)^2}{\rho_G} \tag{24}$$

4.1.2. Design of absorption column

Absorption without chemical reaction

The design of conventional absorption column this work has taken as reference the algorithm of resolution explained in Separation Operations course.

It is assumed that there are several parameters fixed by the industry or laboratory for which it is designed, these, for example, are:

- The component to absorb.
- Inlet gas flow rate.
- Solute concentration at the inlet of the column that it is fixed by the procedure used by the industry.
- Solute concentration outlet of the column, normally fixed by the environmental law.

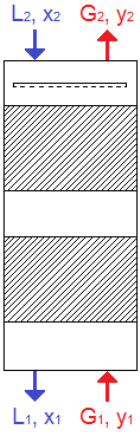

Figure 4.6.: Scheme gas and liquid streams of a conventional packing.

Therefore, to carry out the design of conventional absorption column it will be necessary to fix some parameters.

These parameters are:

The solute to absorb is carbon dioxide.

$G_1 = 0.015$ kg/s = $5.05 \cdot 10^{-4}$ kmol/s

$y_1 = 0.1$

$y_2 = 0.01$

Once these parameters are known proceed to column design.

First of all, it is necessary to choose the absorbing liquid taking into account the solubility of the gas, the price, the danger, if it is not corrosive and inert to the apparatus, etc. For this work it has chosen water as liquid absorbent for the absorption of carbon dioxide. Despite its low absorption capacity and its use is not appropriate as absorbent, it has used in order to compare the liquid and gas phase mass transfer coefficients and to facilitate the design procedure for then compare the absorption conventional column with the Rotating Packed Bed, since this is one of the objectives of the work.

It is considered that the inlet water is free from carbon dioxide, therefore, $x_2 = 0$.

Then it is necessary to fix the liquid flow rate to obtain the outlet concentration desired. To do this it is necessary calculate de minimum inlet flow rate using the equation 7, described in the section 4.1.

To calculate L_{min} is necessary have the equilibrium equation of the carbon dioxide with water to have the value of x_1*. It is assumed that the column works at atmospheric pressure and 25 ºC. In these conditions, the equilibrium equation, using the Henry's law described by equation 6, is:

$$y = 1609.5 \cdot x$$

As aforementioned, the liquid inlet is free of solute, the value of x_2 is 0. With these parameters, using equation 9 it is possible to calculate the value of L_{min}, that is 0.666 kmol/s. As aforementioned, it is necessary that the inlet liquid flow rate should be approximately 1.5 times L_{min}, therefore, L_2 is 0.999 kmol/s and in mass units is 17.93 kg/s.

From this value it is possible to calculate L' using the equation 9. Then with G_1 and equation 8 it is possible to calculate G' and with the latter two variables calculated, it is possible to calculate the outlet liquid composition, x_1, using equation 10. Finally using the variable G' and L' and using again equation 8 and 9, it is possible to obtain the value of G_2 and L_1 respectively. The obtained values of these variables are:

$G' = 4.55 \cdot 10^{-4}$ kmol/(m^2·s) $L' = 0.999$ kmol/(m^2·s)

$G_2 = 4.60 \cdot 10^{-4}$ kmol/s $L_1 = 0.999$ kmol/s

$x_1 = 4.6 \cdot 10^{-5}$

From these results, it is possible to see that the flow rates remained practically constant, therefore the system is diluted. For this reason, the average liquid (L_m) and gas (G_m) flow rate through the column will be used for column's design. These are 0.999 kmol/s and $4.83 \cdot 10^{-4}$ kmol/s respectively.

After calculating these values, it proceeds to choose the type of packing to be used for the column. Due to the fact that absorption of carbon dioxide with water is inefficient, it will be used a packing that provides a large interfacial area to obtain greatest absorption using the minimum water flow rate possible.

The packing used is ceramic Rashig rings of 13 mm because, as aforementioned, it is important that the packing non reaction with neither absorption component. Its characteristics are showed in table A.1 and table A.3. They are found in the appendix 1 and 3.

Wall thickness = 2.4 mm

$C_D = 909$

$\varepsilon = 0.63$

$a_p = 364\ m^2/m^3$

$d_s = 0.01774\ m$

Due to the viscosity of the liquid is smaller than 0.012, the value of $\varphi_{L,W}$ is 0.0325. And consequently de value of ε_L is 0.598

Now that it is already known what packing is used it is necessary to choose a section. This is 3 m^2.

Therefore, the values of L' and G' calculate with the section will be 5.99 $kg/(s \cdot m^2)$ and $4.77 \cdot 10^{-3}$ $kg/(m^2 \cdot s)$ respectively.

With packing characteristics, it is possible to calculate the wetted area thereof using the equation 21 because the liquid is aqueous. For Rashig rings with 13 mm of nominal size and with L' = 6 $kg/(m^3 \cdot s)$, the values of the variables m, n and p are:

$m = 14.69$

$n = 0.01114 \cdot L' + 0.148$

$p = - 0.111$

Thus, the value of a_{AW} is 15.7 m^2/m^3.

Now individual mass transfer coefficients for the gas phase and the liquid phase are determined by equations 12 and 13, taking in to account that the diffusivity of the carbon dioxide on the air and carbon dioxide on water, at 1 atmosphere and 25ºC, are $1.55 \cdot 10^{-5}$ and $2 \cdot 10^{-9}$ respectively.

The gas phase individual coefficient (k_y) is $8.33 \cdot 10^{-5}$ $kmol/(m^2 \cdot s)$ and the liquid phase individual coefficient (k_x) is $2.85 \cdot 10^{-2}$ $kmol/(m^2 \cdot s)$

From individual coefficients values it is possible calculate the overall mass transfer coefficients using equations 14 and 15, thus, the gas phase global coefficient, K_y, is $1.46 \cdot 10^{-5}$ $kmol/(m^2 \cdot s)$ and the liquid phase global coefficient, K_x, is $2.35 \cdot 10^{-2}$ $kmol/(m^2 \cdot s)$. Therefore, global mass transfer coefficients of the gas phase ($K_y a$) and liquid phase ($K_x a$) based on unit volume are $2.30 \cdot 10^{-4}$ $kmol/(m^3 \cdot s)$ and $3.70 \cdot 10^{-1}$ $kmol/(m^3 \cdot s)$.

Once obtained these values it is possible to calculate the volume of the packing using the design equation 17 and 19. The volume is calculating of the gas and liquid phase for then compare that the results to be approximately equal. The results obtained are:

$NUT_G = 5.38$	$NUT_L = 4.43$
$VUT_G = 2.10\ m^3$	$VUT_L = 2.7\ m^3$
$V = 11.30\ m^3$	$V = 11.95\ m^3$

As it can be seen the volumes calculated by the gas phase and the liquid phase are approximately equal.

The pressure drop is not possible to calculated with the figure 4.5 due to it is very low, but to have an approximately idea of its value, it is calculated using equation 24, therefore, the value of the pressure drop per unit of length with dry packing is $1.68 \cdot 10\text{-}2\ N/m^2$.

Absorption with chemical reaction

It has been thought that it would be interesting to design a column where a chemical reaction occurs. For this reason, it has been chosen sodium hydroxide aqueous solution (10%) as an absorbent liquid. To carry out the design of the column the same method has been followed.

The initial parameters which have to start are:

The absorbed solute is carbon dioxide.

$G_1 = 0.015\ kg/s = 5.05 \cdot 10\text{-}4\ kmol/s$

$y_1 = 0.1$

$y_2 = 0.01$

The absorbent liquid reacts instantly with CO_2. The reaction that occurs is:

$$2\ NaOH_{(aq)} + CO_{2(g)} \rightarrow Na_2CO_{3(aq)} + H_2O_{(l)} \qquad \text{(R1)}$$

Due to the reaction, the absorbent liquid does not contain CO_2; therefore, x_1 and x_2 values are 0. As a reaction between the absorbent liquid and the solute, it controls the gas phase. Consequently, only mass transfer coefficient of gas phase is calculated, because $k_G \approx K_G$.

For the column design the liquid flow rate used is 0.666 kmol/s.

The same packed with the same characteristics is also used. The packing used is ceramic Rashig rings of 13 mm.

Wall thickness = 2.4 mm

C_D = 909

ε = 0.63

a_p = 364 m^2/m^3

d_s = 0.01774 m

The characteristics of the packing are showed in table A.1 and table A.3:

Due to the fact that the viscosity of the liquid is smaller than 0.012, the value of $\varphi_{L,W}$ is 0.033, and consequently the value of ε_L is 0.598.

Now that it is already known which packing is used, it is necessary to choose a section. The section of the column will be 3 m^2.

Therefore, the values of L' and G' calculated with the section will be 4.48 $kg/(m^2 \cdot s)$ and $4.77 \cdot 10^{-3}$ $kg/(m^2 \cdot s)$ respectively. As it can be seen, the liquid flow rate with the section is within the range 2 - 6.2 although a low liquid flow rate is used.

With packing characteristics, it is possible to calculate the wetted area thereof using equation 21 because the liquid is aqueous. For Rashig rings with 13 mm of nominal size and with L' = 6 $kg/(m3 \cdot s)$, the values of the variables m, n and p are:

m = 14.69

n = 0.01114·L' + 0.148

p = - 0.111

Thus, the value of a_{AW} is 15.9 m^2/m^3.

Then, individual mass transfer coefficient of the gas phase is determined by equation 12, taking into account that the diffusivity of the carbon dioxide on the air, at 1 atmosphere and 25ºC, is $8.33 \cdot 10^{-5}$ $kmol/(m^2 \cdot s)$.

From the individual coefficient value, it is possible to calculate the global mass transfer coefficient using equation 14, thus, the gas phase global coefficient, as aforementioned, is equal to individual mass transfer coefficient. Therefore, K_g, is also $8.33 \cdot 10^{-5}$ $kmol/(s \cdot m^2)$.

Finally, global mass transfer coefficient of the gas phase ($K_g a$) based on unit volume is $1.33 \cdot 10^{-3}$ $kmol/(s \cdot m^3)$.

Once these values are obtained, it is possible to calculate the volume of the packing using the design equation 17. The volume of the packed is:

NUTG = 2.30

VUTG = 0.36 m^3

V = 0.84 m^3

4.2. MATHEMATICAL MODEL AND DESIGN FOR ROTATING PACKED BED

4.2.1 Mathematical model

The design of a Rotating Packed Bed mainly depends on the relationship of the flow rate and concentration of the inlet and outlet gas and liquid streams, the speed of absorption and mass transfer coefficients, as in the case of the column conventional absorption but also depends on the rotation speed.

For its design, this work has taken as reference the article published by V. Lava Agarwal, from the Chemical Engineering Department of the Indian Institute Technology Kanpur (Lava Agarwal, 2010).

Then it is shown schematically Rotating Packed Bed with its currents:

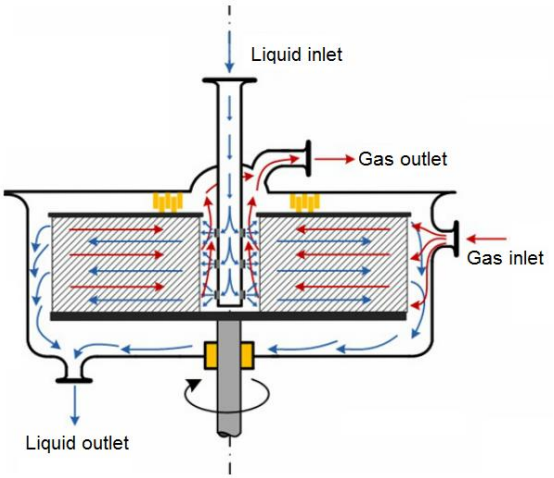

Figure 4.5.: Diagram of a Rotating Packed Bed.

The rotating packed bed consists, as mentioned above, in a rotor that rotates on its axis and is covered by housing. The rotor contains a packing wherein the liquid and gas will contact.

The liquid enters through the pipe located at the top and central part of the rotor and due to shaft rotation, provides a centrifugal acceleration in the liquid so that it is evenly distributed through the packing from the center of the rotor to the periphery and out the bottom of the rotor. The gas enters through the lateral of the housing and it moves from the periphery to toward the center and it leaves the top of the rotor.

As in the same case of conventional column, to obtain a good contact between the liquid and the gas it is important that these have the best distribution through the packing to providing the greater surface area. Therefore, the packing is very important in the design of Rotating Packed bed, just as it had in the conventional column, but in this case, also has a great importance the rotational speed, since it influences on the liquid distribution.

To calculate flow rates mass balances should be made, as in the conventional column, and through them calculate the liquid flow needed to obtain the desired concentration in the outlet. Also the same types of packing can be used.

First of all, it is necessary to decide the rotational speed (rpm), provided that it is reasonable. Instead, the value of the minimum internal radius is necessary to calculate. This is calculated by the following expression:

$$r_i = \left(\frac{G}{\pi \cdot v_{jet} \cdot (1 - f_d)}\right)^{1/2} \cdot \left(\frac{\rho_G \cdot p}{\rho_L}\right)^{1/4} \tag{25}$$

where, the v_{jet} value is between 4 and 5 m/s, the value of f_d is between 1/4 and 1/3 and *p* value is 4.

As this value calculated is the minimum internal radius that can be used, from this value, has to fix a reasonable inner radius.

For calculate the external radius it is necessary to suppose a value for calculate the parameters of the Rotating Packed Bed, once the parameters are calculated is essential to prove that the value is correct using the following equation:

$$r_o = \sqrt{\frac{V}{\pi \cdot h} + r_i^2} \tag{26}$$

To calculate the height of the packing, it is necessary to calculate before the superficial gas velocity inside the Rotating Packed Bed:

$$U_G = \left(\frac{\beta \cdot N_g^a \cdot a_p^b \cdot \mu^c \cdot (\rho_L - \rho_G)^{0.25}}{\left\{ \rho_G^{0.5} + \lambda \cdot \left(\frac{L_v}{\alpha \cdot G_v} \right)^{0.5} \cdot \rho_L^{0.5} \right\}} \right)^2 \quad (27)$$

where, a, b, and c are Sherwood and Wallis correlations and they have the values of 0.43, -0.03 and -0.93 respectively. The Rotating Packed Bed flooding correlations packing parameters, α, β, λ are 1, 130 and 1.51 respectively and Ng is:

$$N_g = \frac{\omega^2 \cdot r_i}{g} \quad (28)$$

After calculating the value of superficial gas velocity in the RPB, already it is possible to calculate the height using the following expression:

$$h = \frac{G_v}{2 \cdot \pi \cdot r_i \cdot U_G} \quad (29)$$

It is important to note that the external radius of Rotating Packed Bed corresponds to the packing height of the conventional column, since it is where the gas and liquid move in the axial direction and the height off RPB corresponds to the radius of the conventional packing column since here the gas and liquid move in the radial direction.

The interfacial wetted area of the packing is calculated using an Onda's equation (M.S. Jassim, 2002)

$$\frac{a_w}{a} = 1 - exp\left[-1.45 \cdot \left(\frac{\sigma_c}{\sigma_L} \right)^{0.75} \cdot \left(\frac{L_w}{a \cdot \mu_L} \right)^{0.1} \cdot \left(\frac{{L_w}^2 \cdot a}{\rho_L^2 \cdot g} \right)^{-0.05} \cdot \left(\frac{{L_w}^2}{\rho_L \cdot \mu_L \cdot a} \right)^{0.2} \right] \quad (30)$$

where, it is important to mention that $g = \omega^2 \cdot r$ and L_w is the mass flow rate (kg/s).

As aforementioned, it is important that the wetted interfacial area should be smaller than the dry area of the packing.

Mass transfer coefficients

To calculate the values of the coefficients of mass transfer in the liquid phase and the gas phase, the correlations obtained by Rajan et al. and Reddy et al. are used respectively.

Coefficient in the liquid phase (Rajan, S. K., 2008):

$$k_L a_e = 0.0733 \cdot Re_L^{0.3547} \cdot Gr_L^{0.2934} \cdot Sc_L^{0.5} \cdot \left(\frac{a_p \cdot D_{BA}}{d_p}\right)^{0.8878} \quad (31)$$

where, the values of the dimensionless numbers are:

$$Re_L = \frac{\rho_L \cdot L_v \cdot d_p}{\mu_L \cdot (2\pi r h)} \quad (31a)$$

$$Gr_L = \frac{d_p^3 \cdot \omega^2 \cdot r \cdot \rho_L^2}{\mu_L^2} \quad (31b)$$

$$Sc_L = \frac{\mu_L}{\rho_L \cdot D_{BA}} \quad (31c)$$

Coefficient in the gas phase (Reddy, K. J., 2006):

$$k_G a_e = 0.00738 \cdot Re_G^{0.976} \cdot Gr_G^{0.132} \cdot Sc_G^{0.333} \cdot \left(\frac{a_p \cdot D_{BC}}{d_p}\right) \quad (32)$$

where, the values of the dimensionless numbers are:

$$Re_G = \frac{\rho_G \cdot G_v \cdot d_p}{\mu_G \cdot (2\pi r h)} \quad (32a)$$

$$Gr_G = \frac{d_p^3 \cdot \omega^2 \cdot r \cdot \rho_G^2}{\mu_G^2} \quad (32b)$$

$$Sc_G = \frac{\mu_G}{\rho_G \cdot D_{BC}} \quad (32c)$$

where, d_p is the effective diameter of packing and t is calculated by the following expression:

$$d_p = 6 \cdot \frac{(1-\varepsilon)}{a_p} \quad (33)$$

After obtaining the values of individual coefficients can be calculated the global coefficients of mass transfer in the same way as have been calculated for conventional column, using the equations 12 and 13.

From these values can be calculated volume RPB using the equation design described above, equations 17 and 19. After calculate the volume of the packing, as aforementioned, it is necessary to prove that the value is correct using the equation 26.

If the outer radius calculated is different from the supposed value will be necessary to suppose another value to the external radio, if the calculated outer radius is equal to the supposed value, then this will be right.

Pressure drop

At the Rotating Packed Bed, as in packed columns, there is also a pressure drop. In this case the pressure drop can be divided into 3 parts: frictional pressure drop, momentum pressure drop and centrifugal pressure drop. The expressions that define these pressures are as follows:

Frictional pressure drop is obtained to considering the packing as a fluidized bed, thus, tacking the Ergun equation and rearranging is obtained the following expression:

$$\Delta P_f = 150 \cdot \frac{\mu_G \cdot (1-\varepsilon)^2}{\varepsilon^3 \cdot d_p^2} \cdot \frac{G}{2\pi h} \cdot \ln\frac{r_o}{r_i} + 1{,}75 \cdot \frac{\rho_G \cdot (1-\varepsilon)}{d_p \cdot \varepsilon^2} \cdot \left(\frac{G}{2\pi h}\right)^2 \cdot \left(\frac{1}{r_i} - \frac{1}{r_o}\right) \quad (34)$$

Momentum pressure drop expression is:

$$\Delta P_m = \frac{1}{2} \cdot \rho_G \cdot \left(\frac{G}{2\pi h \varepsilon}\right)^2 \cdot \left(\frac{1}{r_i^2} - \frac{1}{r_o^2}\right) \quad (35)$$

This last is usually negligible.

Centrifugal pressure drop expression is:

$$\Delta P_C = \frac{1}{2} \cdot \rho_G \cdot A \cdot \omega^2 \cdot (r_o^2 - r_i^2) \quad (36)$$

The total pressure drop of the Rotating Packed Bed is:

$$\Delta P_T = \Delta P_f + \Delta P_C + \Delta P_m \quad (37)$$

4.2.2. Design of Rotating Packed Bed

Absorption without chemical reaction

As the objective is to compare the results obtained in the design of the Rotating Packed bed and the design of the conventional column, to ensure that the values obtained can be compared, the same initial parameters and compositions gas flow rate will be used to design the Rotating Packed bed.

Consequently, the liquid flow rate necessary will be the same value due to, as aforementioned in the section 4.2.1, the liquid flow rate is calculated by the same procedure as in the conventional column. Also the same packing will be used.

These values will be remembered below:

Flow rates characteristics

Gas phase properties	Liquid phase properties
$G_1 = 5.05 \cdot 10^{-4}$ kmol/s = 0.015 kg/s	$L_1 = 0.999$ kmol/s = 17.97 kg/s
$G_2 = 4.60 \cdot 10^{-4}$ kmol/s = $1.36 \cdot 10^{-2}$ kg/s	$L_2 = 0.999$ kmol/s = 17.97 kg/s
$G_m = 4.83 \cdot 10^{-4}$ kmol/s = $1.43 \cdot 10^{-2}$ kg/s	$L_m = 0.999$ kmol/s = 17.97 kg/s
$y_1 = 0.1$	$x_1 = 4.6 \cdot 10^{-5}$
$y_2 = 0.01$	$x_2 = 0$

The solute to absorb is also carbon dioxide and the absorbing liquid used is water.

Packing characteristics: Ceramic Rashig rings, 13mm

Wall thickness = 2.4 mm

$\varepsilon = 0.63$

$a_p = 364$ m^2/m^3

The effective diameter of packing is $6.10 \cdot 10^{-3}$m.

First of all, it is necessary to fix a rotation speed that is reasonable, according to articles read the value of this could be 1000 rpm.

Then, it is important to calculate the internal radius, since individual coefficients material depend on it. This is calculated by equation 25, and the value obtained is $8.56 \cdot 10^{-3}$ m. This is the minimum inner radius that can be used, but as within the radius must have the liquid distributor, has to fix a reasonable inner radius. Its value is 0.04 m.

As aforementioned, it is necessary to suppose an outer radius, the value of this is 1.22 m.

Besides the radius, to calculate the individual coefficients is also necessary calculate the height of the packing, but before it is necessary calculate the superficial gas velocity inside the Rotating Packed Bed with equation 27. The superficial gas velocity is $8.95 \cdot 10^{-2}$ m/s.

After that it is possible to calculate the packing height using equation 29, this is 0.52 m.

The interfacial wetted area of the packing is calculated using an Onda's expression aforementioned, equation 30. It is important to mention, that the ceramic critical surface tension

is 0.061 N/m and the solution critical surface tension is 0.0726 N/m, therefore the interfacial wetted area of the packing is 360 m^2/m^3.

Now individual mass transfer coefficients for the gas phase and the liquid phase are determined by the correlations obtained by Rajan et al. and Reddy et al., these are showed with equations 31 and 32.

The diffusivity of the carbon dioxide on the air and carbon dioxide on water, at 1 atmosphere and 25ºC, are, as mentioned before, $1.55 \cdot 10^{-5}$ and $2 \cdot 10^{-9}$ respectively.

The value of the coefficient in the liquid phase (Rajan, S. K., 2008) is 1.21 s^{-1} and this in the gas phase (Reddy, K. J., 2006) is 0.21 s^{-1}.

Once obtained the values of individual coefficients can be calculated the global coefficients of mass transfer in the same way as have been calculated for conventional column, using equations 14, for the gas phase, and 15, for the gas phase, these values are $1.07 \cdot 10^{-3}$ $kmol/(m^3 \cdot s)$ and 1.72 $kmol/(m^3 \cdot s)$ respectively.

From these values can be calculated volume of the Rotating Packed Bed using equation design described above, equation 17 and 19. For the Rotating Packed bed it has also been calculated the volume of the gas and liquid phase in order to compare them and see that they are approximately equal. The results obtained are:

$NUT_G = 5.38$	$NUT_L = 4.43$
$VUT_G = 0.45\ m^3$	$VUT_L = 0.58\ m^3$
$V = 2.43\ m^3$	$V = 2.57\ m^3$

As it can be seen the volumes calculated by the gas phase and the liquid phase are approximately equal. The total pressure drop, as aforementioned before, is the sum of the frictional, momentum and centrifugal pressure drop, the obtained values are 0.6 Pa, 0.01 Pa and $5.23 \cdot 10^4$ Pa respectively. As it is shown that the momentum pressure drop is the lowest and therefore can be neglected, but the frictional pressure drop is also low, this may be because the gas flow rate is very low. Finally, the total pressure drop through the column is 52.30 kPa.

Absorption with chemical reaction

As in the same case as in the conventional column, the Rotating Packed Bed was also designed for the case that it occurs a chemical reaction. For that reason, it can also be used as

aqueous absorbent of sodium hydroxide. Therefore, both the same initial parameters and gas flow rate compositions will be used.

As mentioned earlier, as there is no chemical reaction between the absorbent liquid and solute, the solute concentration in the liquid stream will be 0.

The fixed flow rates characteristics are showed below:

Gas phase properties	Liquid phase properties
$G_1 = 5.05 \cdot 10^{-4}$ kmol/s = 0.015 kg/s	$L_1 = 0.999$ kmol/s = 23.46 kg/s
$G_2 = 4.60 \cdot 10^{-4}$ kmol/s = $1.36 \cdot 10^{-2}$ kg/s	$L_2 = 0.999$ kmol/s = 17.97 kg/s
$G_m = 4.83 \cdot 10^{-4}$ kmol/s = $1.43 \cdot 10^{-2}$ kg/s	$L_m = 0.999$ kmol/s = 17.97 kg/s
$y_1 = 0.1$	$x_1 = 0$
$y_2 = 0.01$	$x_2 = 0$

As in the same case as in the other designed process to compare both types of methods studied in this work, the same packing is used:

Packing characteristics: Ceramic Rashig rings, 13mm

Wall thickness = 2.4 mm

$\varepsilon = 0.63$

$a_p = 364$ m^2/m^3

The effective diameter of packing is $6.10 \cdot 10^{-3}$m.

First of all, it is necessary to fix a rotation speed that is reasonable. According to read articles, this value could be 1000 rpm.

Then, it is necessary to calculate the inner radius. It will be calculated using equation 25. The inner radius obtained is $8.30 \cdot 10^{-3}$ m. This, as aforementioned, is the minimum inner radius that can be used, thus the internal radius used has a value of 0.04 m.

As mentioned before, it is necessary to suppose an outer radius, and then it is also necessary to verify that it is correct. This value is 0.73 m.

Besides the radius, in order to calculate the individual coefficient, it is also necessary to calculate the height of the packing. But, before that, it is necessary to calculate the superficial gas velocity using equation 27. The superficial gas velocity is $1.26 \cdot 10^{-1}$ m/s.

After that, it is possible to calculate the packing height using equation 29. This is 0.37 m.

The interfacial wetted area of the packing is calculated using an aforementioned Onda's expression: equation 30. It is necessary to remember that that the ceramic critical surface tension is 0.061 N/m and the solution critical surface tension is considered approximately at the rang of pure water, because the solution has a low concentration of NaOH. This value is 0.0726 N/m. Therefore, the interfacial wetted area of the packing is 319.3 m^2/m^3.

Thus, individual mass transfer coefficient for the gas phase is determined by the correlation obtained by Rajan et al. this is showed with equation 32.

The diffusivity of the carbon dioxide on the air, at 1 atmosphere and 25ºC, is, as mentioned before, $1.55 \cdot 10^{-5}$ m/s.

The value of the coefficient in the gas phase (Reddy, K. J., 2006) is 0.41 s^{-1}.

Once the value of individual coefficient is obtained, the mass transfer global coefficient can be calculated using equation 14, that, as aforementioned, has the same value of the individual mass transfer coefficient because a chemical reaction occurs, and consequently the gas phase controls. This value is $1.82 \cdot 10^{-3}$ kmol/($m^3 \cdot$s).

From these values, the volume of the Rotating Packed Bed can be calculated using the design equation described above, equation 17. The results are:

$$NUT_G = 2.30$$
$$VUT_G = 0.26\ m^3$$
$$V = 0.61\ m^3$$

5. COMPARISON OF DEVICES

This section compares the obtained results in the design of the conventional column and in the Rotating Packed Bed calculated in the previous section. Between the all parameters calculated the most interesting are showed in the table 5.1.

	Conventional column		Rotating Packed Bed	
	Without chemical reaction	With chemical reaction	Without chemical reaction	With chemical reaction
K_La, kmol/(m^3·s)	$3.7\cdot10^{-1}$	-	1.72	-
K_Ga, kmol/(m^3·s)	$2.3\cdot10^{-4}$	$1.33\cdot10^{-3}$	$1.07\cdot10^{-3}$	$1.82\cdot10^{-3}$
V_L, m^3	11.30	0.84	2.43	0.61
a_w, m^2/m^3	15.73	15.909	360	319.3

Table 5.1.: Characteristics comparison of the Rotating Packed Bed and conventional column.

Comparison between Rotating Packed Bed and conventional column

If the coefficients of mass transfer per unit of volume are compared, the coefficients obtained in the rotating packed bed are higher that the coefficients obtained in the conventional column, as it was expected. Surely due to, as it can be observed, the liquid distribution in the packing is greater in Rotating Packed Bed than in conventional column using the same packing and this, as aforementioned, enhancement the mass transfer because the liquid has a large superficial area.

Nevertheless, the difference between the mass coefficients is very small. This explains why the authors as David B. Todd, Fen Guo et al., Ji et al., Guang-Wen and Yong L. C. et al. have conducted a series of experiments, in which, as explained before, they have modified the contact type between the fluids, in order to try promote absorption process in the Rotating Packed Bed. However, these experiments have not made improvements.

On the other hand, it is interesting to note that for the same amount absorbed, the volume of the Rotating Packed Bed is much lower than the volume of the conventional absorption column, this represents an advantage because the volume occupied by the device can be a limitation.

Comparison between the process with chemical reaction and without chemical reaction

If results carried out in both the Rotating Packed Bed and in the conventional column are compared, one can say that the absorption process with chemical reaction favors the coefficients of mass transfer and it reduces the volume of the device. Nevertheless, the specific wetting area, in the Rotating Packed Bed, decreases. This may be because the absorbent liquid used in the absorption with chemical reaction has a higher viscosity than the one used in the absorption without chemical reaction. Consequently, it causes a worse liquid distribution through the packing. In the case of the conventional column, the wetted area is approximately equal due to it does not depends on the liquid viscosity.

Apart of this, it is necessary to emphasize that the obtained results in the Rotating for carrying out an absorption with or without chemical reaction, are similar.

The results obtained in the Rotating Packed Bed not show great improvement absorption compared to conventional column, for this reason, it has decided to change the rotational speed and size of the packing to see how it effects on the characteristics of the device.

First, it has been studied as global mass transfer coefficients varying in the gas phase and liquid phase, the outer radius, with an inner radius of 0.04 m, the wetted area and finally, the height and volume of the packing. The obtained results are shown below:

ω (rpm)	ω (rad/s)	$K_G a$ kmol/(m³·s)	$K_L a$ kmol/(m³·s)	r_o (m)	h (m)	V (m³)	ΔP_c (kPa)
600	62.83	3.26E-04	0.52	1.43	1.24	8.44	25.9
800	83.78	6.31E-04	1.01	1.32	0.76	4.36	39.2
1000	104.72	1.07E-03	1.72	1.22	0.52	2.57	52.3
1200	125.66	1.66E-03	2.68	1.15	0.38	1.65	66.9
1400	146.61	2.43E-03	3.91	1.09	0.29	1.13	81.8
1600	167.55	3.39E-03	5.46	1.03	0.23	0.81	95.4
1800	188.50	4.57E-03	7.35	0.98	0.19	0.60	109.3

Table 5.2.: variation of the characteristics of RPB front different-rotation speeds.

As shown in the results, by increasing the rotational speed, mass transfer between the liquid and the gas is favored and equipment size is reduced. This may be because increasing the

rotational speed, the wetted area is favored, so there is a better liquid distribution and better contact between the liquid and gas phase.

On the one hand, these results are very positive and show that with less volume it is possible to obtain greater mass transfer between the liquid and the gas. But, on the other hand, it can be negative because a higher rotational speed involves a higher consumption energy, the equipment is more complex and consequently more expensive to manufacture, and may be a faster machine wear occurs.

Besides this, it is shown that with increasing rotational speed also increases the centrifugal pressure drop, as was expected.

Then it is also considered interesting to study how influences the packing nominal size, which fixed the porosity and specific area of the packing, in the characteristics of the Rotating Packed Bed. For this purpose, ceramic Raschig rings are used with different nominal sizes. The following table shows the obtained results:

Nominal size, mm	**ε**	**a_p (m^2/m^3)**	**a_w (m^2/m^3)**	**K_Ga kmol/(m^3·s)**	**K_La kmol/(m^3·s)**	**r_o (m)**	**h (m)**	**V (m^3)**
6	0.73	787	765.2	$4.14 \cdot 10^{-4}$	0.67	0.96	2.16	6.65
9.5	0.68	508	499.5	$7.04 \cdot 10^{-4}$	1.13	1.11	0.96	3.90
13	0.63	364	360.0	$1.07 \cdot 10^{-3}$	1.72	1.23	0.52	2.57
16	0.68	328	324.9	$1.13 \cdot 10^{-3}$	1.82	1.31	0.42	2.43
1	0.73	262	260.2	$1.35 \cdot 10^{-3}$	2.18	1.48	0.28	2.03
25	0.73	190	189.2	$1.95 \cdot 10^{-3}$	3.13	1.66	0.15	1.41
32	0.74	148	147.6	$2.55 \cdot 10^{-3}$	4.11	1.83	0.10	1.08

Table 5.3.: variation if the characteristics of RPB front different packing nominal size.

As it can be seen from the above results, for the same type of packing, but with different nominal sizes, some unexpected results are obtained. As the nominal size increases, the specific area of the packing decreases and as a result, the mass transfer should decrease, as this is favored by the contact area between liquid and gas. However, instead of decreasing, it increases. It can be caused by the fact that the wetted specific area gets closer to the packing

superficial area when the packing has a bigger nominal size. Theoretically, it means that there is a better liquid distribution and therefore the mass transfer increases.

Comparing the results obtained with the experimental results obtained by other authors, it can be seen that the same results are obtained for different types of experiments. Hereafter some examples are showed:

Yong Luo et al. studied the effective interfacial area in the Rotating Packed Bed and obtain the following results (Yong, L., 2012):

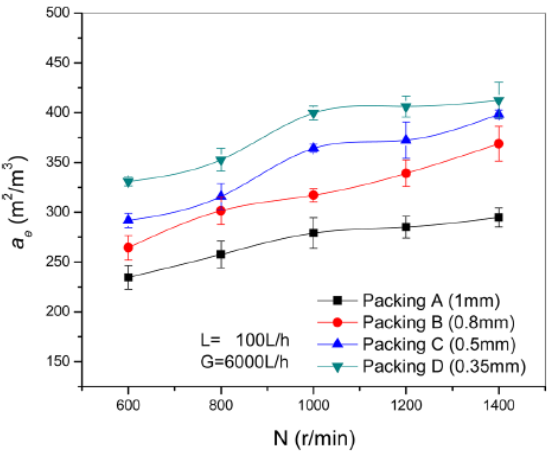

Figure 5.1.: Effects of the rotational speed and nominal size on specific area.

As it can be observed, both by increasing the nominal size of the packing as well as increasing the rotation speed of the device, there is an increase of the specific wetting area, just as the results obtained in this work.

Liang-Liang Zhang et al. studied the absorption of carbon dioxide with ionic liquid. The results that they obtained about the evolution of the individual mass transfer coefficient of the liquid phase for different rotational velocity are (Liang-Liang, Z., 2011):

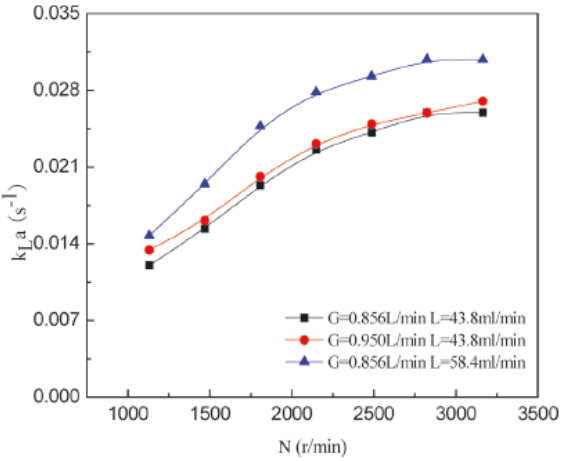

Figure 5.2.: Effect of the rotational speed on the mass transfer coefficient.

As it can be observed the effect of the rotational speed also influenced on mass transfer coefficient, this value increases with the rotational speed, as in the results obtained in this work.

Finally, through studies by A. Chandra, has also been found that the pressure drop due to the centrifugal force increases with the rotational speed, just as shown in the results obtained in this work. The results are shown in the following graphic:

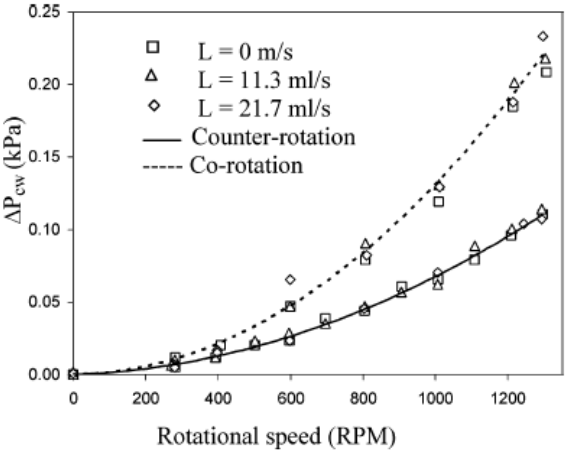

Figure 5.3.: Effect of the rotational speed on the centrifugal pressure drop.

Therefore, it is verified that the theoretical results obtained in this study are in the same range with the experimental results obtained by the several authors that have been studied.

6. CONCLUSIONS

The Higee technology has been developed to enhance various industrial processes such as adsorption, desorption, distillation, extraction and stripping among others.

This paper has studied the Higee technology applied to absorption in order to compare its performance with the one observed in the conventional absorption column. According to the literature study, the expected improvements are:

- The use of this technology improves process performance, allowing a better distribution of the liquid through the packing and enhancing the mass transfer between the gas and the liquid.

- Decreases the size of the equipment and, consequently, its weight.

- For the same mass transfer rate it uses less absorbent.

- Faster absorption due to its greater interaction between the liquid and gas.

- Reduces capital costs and energy consumption.

Although Best available techniques Reference document (BREF) explains that the Higee technology is a very promising technology in the field of Technique for Manufacturing of Organic Fine Chemicals, it has not achieved the expected results. However, the theoretical results obtained in this study correspond to the experimental results obtained by the studied authors. Therefore, it is concluded that:

- The values of the mass transfer coefficients obtained using Higee technology are approximately at the same range to those obtained with a conventional column.

- Although the process is faster, more energy is consumed because the device must be rotated.

- The design of the device is more complex and more expensive.

- The fact that the apparatus rotates on its shaft, implies a faster wear thereof.

Therefore, it is concluded that the use of Higee technology is not highly recommended because its advantages compared to the ones observed on the conventional technology are not remarkable.

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NOMENCLATURE

a, b, c: RPB flooding correlation fitting parameters

A: centrifugal pressure drop proportionality constant.

a_{AW}: specific area, [m^2/m^3]

a_p: specific surface area of the packing, [m^2/m^3]

D_{AC}: diffusion coefficient of liquid, [m/s]

D_{BC}: diffusion coefficient of gas, [m/s]

d_p: effective diameter of packing, [m]

d_s: diameter of the sphere, [m]

f_d: fraction of cross-sectional area of RPB eye occupied by distributor

g: gravitational force, [m/s^2]

G‘: mass gas flow rate per unit of section, [$kg/(m^2 \cdot s)$]

G_1: inlet gas flow rate, [kmol/s]

G_2: outlet gas flow rate, [kmol/s]

G_m: mass gas flow rate, [kg/s]

Gr: Grashof number

G_v: volumetric gas flow rate, [m^3/s]

H: Henry’s constant

h: packing height, [m]

k_G a: gas-side volumetric mass transfer coefficient, [s^{-1}]

K_G a: global gas volumetric mass transfer coefficient, [$kmol/(m^3 \cdot s)$]

K_L a: global liquid volumetric mass transfer coefficient, [$kmol/(m^3 \cdot s)$]

k_L a: liquid-side volumetric mass transfer coefficient, [s^{-1}]

L’: mass liquid flow rate per unit of section, [$kg/(m^2 \cdot s)$]

L_1: outlet liquid flow rate, [kmol/s]

L_2: inlet liquid flow rate, [kmol/s]

L_m: mass liquid flow rate, [kg/s]

L_v: volumetric liquid flow rate, [m^3/s]

N_g: ratio of centrifugal to gravitational acceleration

P: pressure, [Pa]

P_B: partial pressure, [Pa]

P_{VB}: vapor pressure of the component B, [Pa]

Re: Reynolds

r_i: inner radius, [m]

r_o: outer radius, [m]

S: section, [m^2]

S_c: Schmidt number

U_G: gas superficial velocity, [m/s]

V: volume, [m^3]

v_{jet}: liquid distributor jet velocity, [m/s]

x*: component liquid-phase mole fraction in equilibrium with gas phase

x: component mole fraction in liquid phase

y*: component gas-phase mole fraction in equilibrium with liquid phase

y: component mole fraction in gas phase

ΔP_c: centrifugal pressure drop, [N/m^2]

ΔP_f: frictional pressure drop, [N/m^2]

ΔP_m: momentum pressure drop, [N/m^2]

ΔP_T: total pressure drop, [N/m^2]

ε_L: porosity of the packing

λ, α, β: RPB flooding correlation fitting parameters

μ_G: gas viscosity, [kg/(m·s)]

μ_L liquid viscosity, [kg/(m·s)]

ρ_G: gas density, [kg/m^3]

ρ_L: liquid density, [kg/m^3]

ρ_M: molar density, [kmol/m^3]

σ_c: critical surface tesion for ceramic, [N/m]

σ_L: critical surface tension for solution, [N/m]

$\varphi_{L,w}$: fluid retention

ω: RPB rotational speed, [rad/s]

APPENDICES

APPENDIX 1: PACKING CHARACTERISTICS

			Nominal size (mm)									
Packing	Material	Characteristics	6	9.5	13	16	19	25	32	38	50	76
Rashig ring	Ceramic	C_D			909	749	457	301		101,8	135,6	
		ε	0.73	0.68	0.63	0.68	0.73	0.73	0.74	0.71	0.74	0.78
		a_p (m^2/m^3)	787	508	364	328	262	190	148	125	92	62
	Metal	C_D			688	431	485	304		172.9	133.5	
		ε			0.73		0.78	0.85	0.87	0.90	0.92	0.95
		a_p (m^2/m^3)			387		236	286	162	135	103	68
Pall rings	Plastic	C_D				297		105.2		61.8	47.5	
		ε				0.87		0.90		0.91	0.92	
		a_p (m^2/m^3)				341		206		128	102	
	Metal	C_D				133.4		95.5		56.6	36.5	
		ε				0.93		0.94		0.95	0.96	
		a_p (m^2/m^3)				341		206		128	102	
Berl saddles	Ceramic	C_D			508		295	184				
		ε	0.60		0.63		0.66			0.75	0.72	
		a_p (m^2/m^3)	899		466		269			144	105	
Intalox saddles	Ceramic	C_D			399		256	241.5		96.2	71.3	40.6
		ε	0.75		0.78		0.77	0.775		0.82	0.79	
		a_p (m^2/m^3)	984		623		335	256		195	118	
	Plastic	C_D					96.7			56.5	30.1	
		ε					0.91			0.93	0.94	
		a_p (m^2/m^3)					207			108	89	
Super intalox	Ceramic	C_D					123			63.3		
		ε					0.79			0.81		
		a_p (m^2/m^3)					253			105		
	Plastic	C_D					79.5			53.5	30.1	
		ε					0.90			0.93	0.94	
		a_p (m^2/m^3)					207			104	89	

Table A.1.: Characteristics of different types of packing.

APPENDIX 2: INTERFACIAL AREA FOR AQUEOUS LIQUIDS

Packing	Nominal size, mm	Range L', kg/(m²·s)	m	n	p
Rashig rings	13	0.68- 2.0	28.1	0.2323 *L'* – 0.30	-1.04
		2.0 – 6.1	14.69	0.01114 *L'* + 0.148	-0.111
	25	0.68- 2.0	34.42	0	0.552
		2.0 – 6.1	68.2	0.0389 *L'* – 0.0793	-0.47
	38	0.68- 2.0	36.5	0.0498 *L'* – 0.1013	0.274
		2.0 – 6.1	40.11	0.01091 *L'* – 0.022	0.140
	50	0.68- 2.0	31.52	0	0.481
		2.0 – 6.1	34.03	0	0.362
Berl saddles	13	0.68- 2.0	26.28	0.0529	0.761
		2.0 – 6.1	25.61	0.0529	0.170
	25	0.68- 2.0	52.14	0.0506 *L'* – 0.1029	0
		2.0 – 6.1	73.0	0.0319 *L'* – 0.0630	– 0.359
	38	0.68- 2.0	40.6	- 0.0508	0.455
		2.0 – 6.1	62.4	0.0249 *L'* – 0.0996	– 0.1355

Table A.2.: Constants values for calculating interfacial area for aqueous liquids.

APPENDIX 3: FLUID RETENTION

Packing: Ceramic Rashig rings

Nominal size, mm	d_s, m	φ_{Ls}	μ_L, kg/(m·s)	Water,ordinary temperatures
13	0.01774	$\frac{0.0486 \cdot \mu_L^{0.02} \cdot \sigma^{0.99}}{d_s^{1.21} \cdot \rho_L^{0.37}}$	<0.012	$\beta = 1.508 \cdot d_s^{0.376}$ $\varphi_{L,W} = \frac{2.47 \cdot 10^{-4}}{d_s^{1.21}}$
25	0.0356			
38	0.0530		>0.012	$\varphi_{L,W} = \frac{2.09 \cdot 10^{-6} \cdot (737.5 \cdot L')^{\beta}}{d_s^2}$
50	0.0725			
Packing: Carbon Rashig rings				
25	0.01301	$\frac{0.0237 \cdot \mu_L^{0.02} \cdot \sigma^{0.23}}{d_s^{1.21} \cdot \rho_L^{0.37}}$	<0.012	$\beta = 1.104 \cdot d_s^{0.376}$ $\varphi_{L,W} = \frac{5.94 \cdot 10^{-4}}{d_s^{1.21}}$
38	0.0543			
50	0.0716		>0.012	$\varphi_{L,W} = \frac{7.34 \cdot 10^{-6} \cdot (737.5 \cdot L')^{\beta}}{d_s^2}$
Packing: Ceramic Berl saddles				
13	0.31622	$\frac{4.23 \cdot 10^{-3} \cdot \mu_L^{0.04} \cdot \sigma^{0.55}}{d_s^{1.56} \cdot \rho_L^{0.37}}$	<0.020	$\beta = 1.508 \cdot d_s^{0.376}$ $\varphi_{L,W} = \frac{5.014 \cdot 10^{-5}}{d_s^{1.56}}$
25	0.0320			
38	0.0472		>0.020	$\varphi_{L,W} = \frac{2.32 \cdot 10^{-6} \cdot (737.5 \cdot L')^{\beta}}{d_s^2}$

Table A.3.: Parameters to calculate fluid retention in packing columns.